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FULL SCALE PARALLEL ACTIVATED SLUDGE PROCESS EVALUATION

Erwin D. Toerber

National Environmental Research Center
Cincinnati, Ohio

November 1972

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Full Scale Parallel Activated Sludge Process Evaluation

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16. Abstract

A comparison was made between parallel activated sludge systems operating under completely-mixed and plug-flow modes.

Initially, a rhodamine dye tracer study was conducted to determine conditions necessary to achieve the two operational modes.

The completely-mixed system was operated at 5 constant detention times ranging from 3 hours to 1 hour. The break in treatment efficiency (a marked drop below 90% removal of soluble BOD_5) occurred between 1 and 2 hours.

No marked difference in treatment efficiency was found between the two modes during 5 months of the parallel operation. Two methods of solids production analysis were applied to the data. A series of oxygen uptake tests were performed to evaluate air requirements.

Finally, a set of shock loads were applied in parallel and separately to each mode. The complete-mix system did show an advantage over plug-flow under shock load conditions at a short detention time (1.5 hours).

A 500 gpd completely-mixed pilot plant was run in parallel with the full scale system for 4 months. It was most successful in duplicating the full scale organic removal efficiency. (EPA abstract)

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FULL SCALE PARALLEL ACTIVATED SLUDGE PROCESS EVALUATION

By

Erwin D. Toerber

Project 17050 ENM

Project Officer

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SECTION I

CONCLUSIONS

Mixing Study

1. By proper placement of diffusers, use of transverse flow, and maintenance of the stated minimum air flow per unit tank volume, it is possible to achieve completely-mixed conditions with a mean residence time very close to the nominal.

2. To achieve conditions approaching plug-flow an aeration tank system using spiral aeration and a length:width ratio of 18.4:1 was found to be adequate.

Constant Detention Time Study

1. For both completely-mixed and plug-flow under normal conditions ($F/M = 0.2 - 1.0$ lb BOD_5/lb MLVSS), no marked break (a drop below 90% removal of soluble BOD_5) in treatment efficiency was experienced until the nominal aeration detention time fell below 1 - 2 hours.

Parallel Operation

1. Under normal loading conditions ($T = 3.1 - 3.6$ hrs, $F/M = .27 - .82$ lb BOD_5 /lb MLVSS, 25-55 lb $BOD_5/1000\ ft^3$), no appreciable difference could be found between CM and PF on the basis of BOD or COD removal efficiency from primary effluent to final effluent.

2. Two methods of solids production analysis were applied to approximate solids production from these models, however, a large amount of scatter was experienced in the solids data.

3. A series of oxygen uptake tests were performed on the mixed liquor from each system. From these tests it was possible to determine rate constants to predict oxygen requirements using an equation of the form $\frac{do}{dt} = K_sF + K_eM_a$.

   Average values for the constants were determined to be $K_s = 2.6\ hrs^{-1}$ and $K_e = .026\ hrs^{-1}$.

4. Final settling performance was evaluated and it was found that to maintain an effluent SS of 25 mg/l, loadings should not exceed the following:

Overflow Rate = 700 gpd/ft^2
(excluding recirculation)

Solids Loading = 25 lb/day/ft^2
(based on total mixed liquor flow)

Shock Load Studies

1. Both CM and PF reacted very similarly to an organic shock load at a detention time of 2.5 hrs.
2. At a reduced detention time of 1.5 hrs the CM system did show superiority in mass total BOD$_5$ removal (73% for PF compared to 84% for CM).
3. During all the shock loads the CM system demonstrated a less marked drop in aeration tank DO than did PF.

Pilot Plant Studies

1. The pilot plant was found to be reliable as a design tool (on domestic waste) for predicting the following:
   a) BOD removal efficiencies and removal rate constants
   b) Oxygen uptake and consequently air requirements
2. The reliability of the plant for predicting solids production is very questionable.

Design Analysis

1. A set of reliable design parameters were developed from this study, and an economic comparison of a CM plant designed on this basis versus one designed using state standards showed a 2 to 1 factor in favor of the former.
2. The cost analysis outlined in No. 1 was also compared with the EPA economic computer program. The EPA program yielded a total cost of approximately 80% of that determined from costs estimated for a plant designed according to the parameters developed from this report.

SECTION II

RECOMMENDATIONS

Though extensive mixing studies were conducted, it was not until the end of this phase that an optimum design for a CM aeration tank was determined. Additional tracer tests to further verify the findings would be advisable.

The shock load studies showed some advantage in CM at high organic loads and short detention times. It would be good to apply shock loads of longer duration (3-6 hrs) to see if a greater difference in system performance would result.

Much difficulty was experienced in developing a reliable solids prediction equation for either system. More research is needed in relation to sludge flow and concentration measurements and the analysis procedures for such data.

For other researchers looking for small differences in treatment efficiency of biological waste treatment units, it was found that the sampling method is all important. The most reliable samples are grab samples taken on a time dependent basis, e.g., the shock load studies in this report.

SECTION III

INTRODUCTION

Background

The purpose of this study was to compare the performance of a completely-mixed (CM) activated sludge system with a plug flow (PF) system under parallel operating conditions. The original plan also considered a comparison of CM with contact stabilization and step aeration if significant performance differences could be detected under the existing loading conditions.

The project originated when the City of Freeport, Ill. (population 30,000) Water & Sewer Commission expanded its wastewater treatment plant to include secondary treatment of the activated sludge type in 1968-69. The design was to be based on the principles of CM developed to date. However, only sufficient capacity as measured by Illinois EPA standards would be provided for present flow. It was believed that the true capability of the plant would be adequate to handle design future and it was desired to establish that fact before committing additional funds for capacity not actually required.

The Illinois Sanitary Water Board (now Illinois Environmental Protection Agency) approved of this design approach on a trial basis providing that adequate flexibility be included to allow operation under PF and step aeration conditions also.

A schematic of the activated sludge system as built is shown in Figure 1. The plant was constructed so that it could be operated as two entirely separate systems using different flow patterns, mixing regimes, air rates, and return sludge and wasting rates.

The demonstration grant (17050 ENM) awarded by the Environmental Protection Agency, Office of Water Programs, was then to provide for Freeport to utilize the design flexibility of the plant for a full scale activated sludge parallel process evaluation.

Since much research has been conducted on pilot plant and lab-scale concerning CM operation, it was decided to incorporate into the program a 500 gpd pilot plant operating simultaneously with the parallel full scale plants.

Finally, on the basis of the data collected, an analysis

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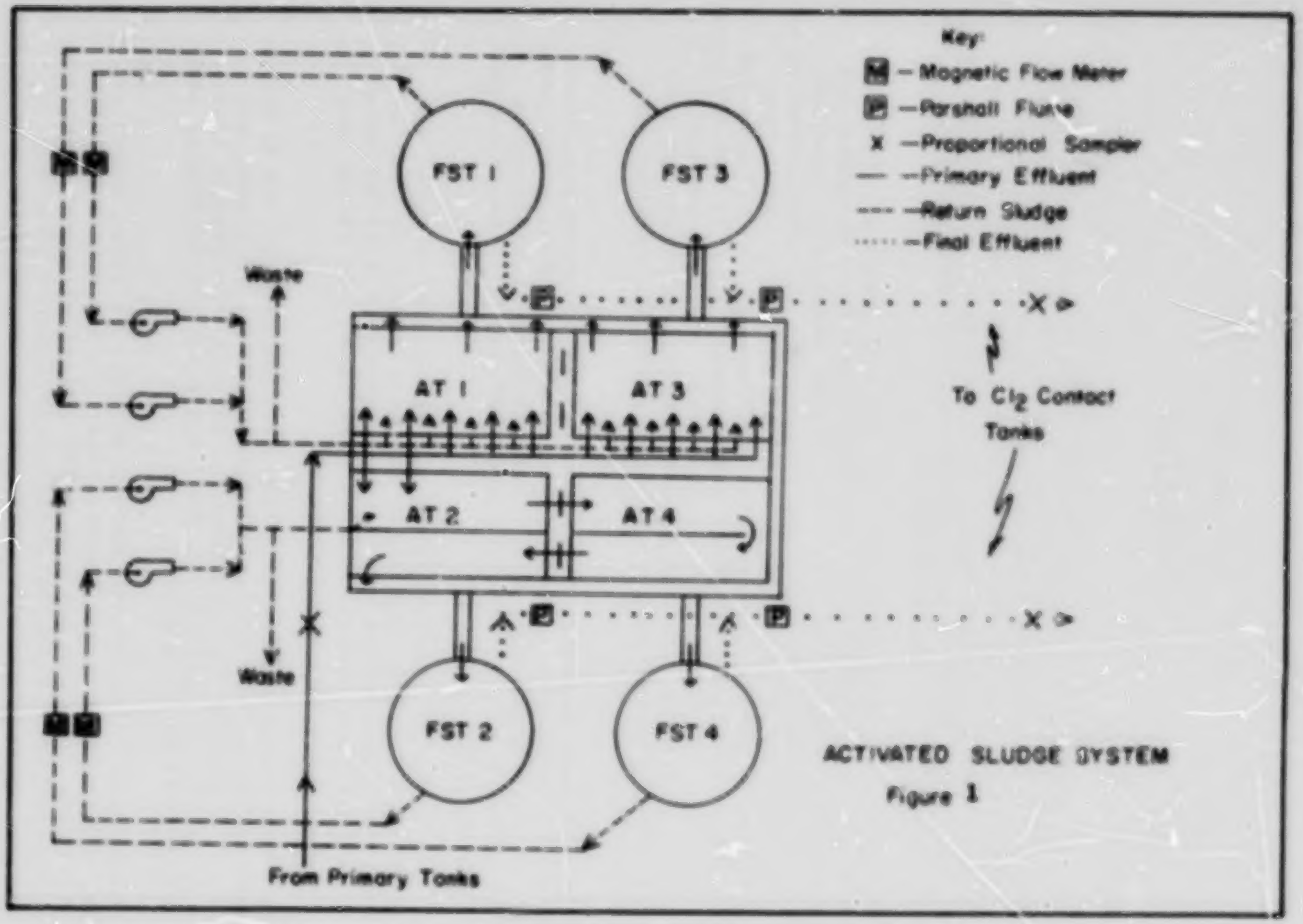

Key:
▣ —Magnetic Flow Meter
🄿 —Parshall Flume
X —Proportional Sampler
—— —Primary Effluent
- - - —Return Sludge
······ —Final Effluent
FST 1
FST 3
FST 2
FST 4
AT 1
AT 3
AT 2
AT 4
Waste
Waste
To Cl2 Contact Tanks
From Primary Tanks

ACTIVATED SLUDGE SYSTEM

Figure 1

of plant design and economic factors would be made. The ultimate objective of the project then being to provide significant information to improve future plant designs.

To carry out the study, a technical committee was formed. The committee met on a quarterly basis to evaluate the progress of the study and to plan the next quarter's work. The committee consisted of the following persons:

Technical Committee Members	Responsibilities
Erwin D. Toerber Freeport Water & Sewer Commission	Project Engineer
Lawrence M. Madden Water & Sewer Commission	Project Manager
H.S. Smith University of Idaho	Research Consultant
Wayne L. Paulson University of Iowa	Research Consultant
Charles Swanson & Richard Eilers	First Mr. Swanson & then Mr. Eilers served as EPA Project Officer
Fred VanKirk Consoer, Townsend & Associates	Design of plant & analysis of data to apply to design recommendations
Robert Bella Consoer, Townsend & Associates	Analysis of data to apply to design recommendations

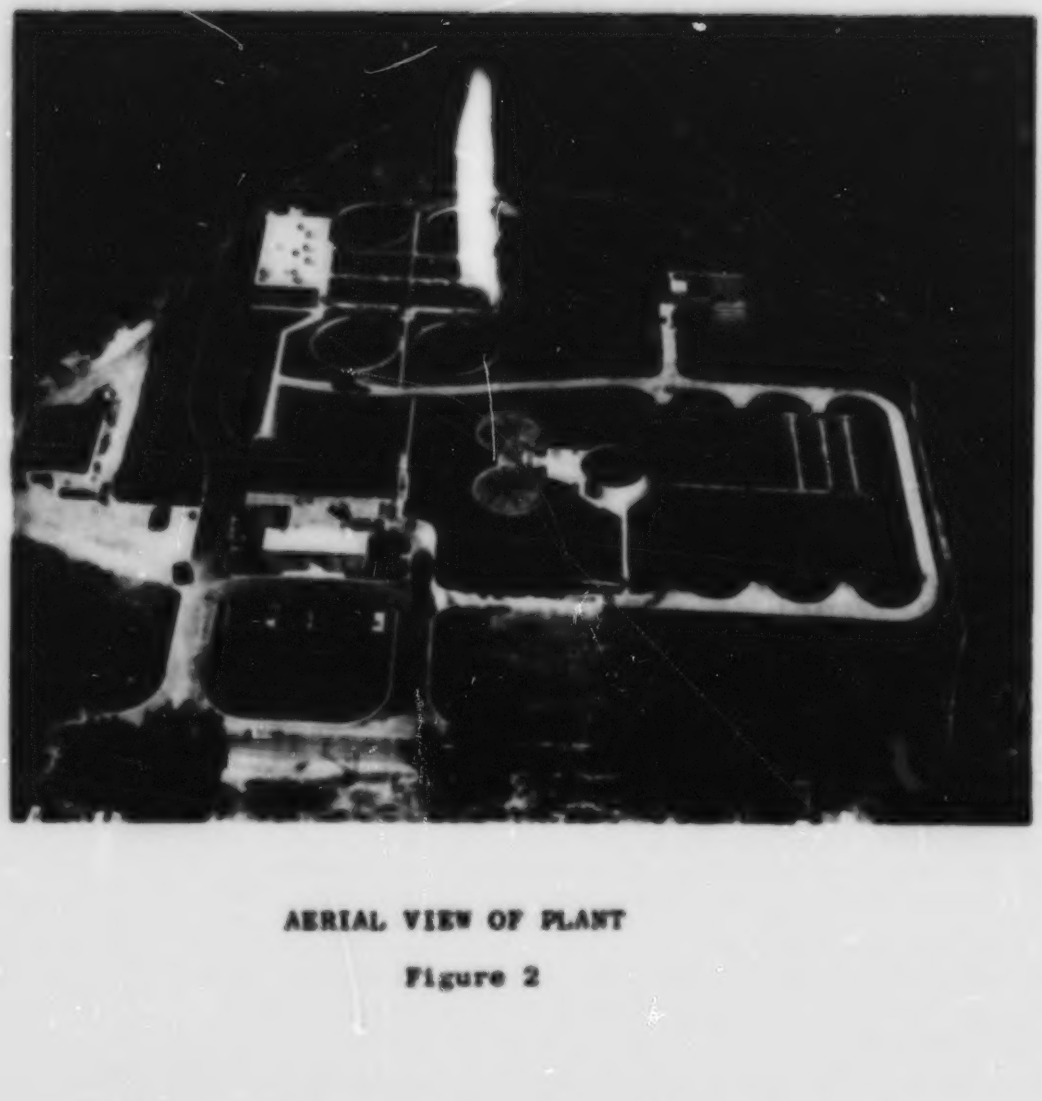
The study was divided into five phases. Each phase and its duration are listed below:

1. Aeration Tank Mixing Study	May-Sept. 1970
2. CM Constant Detention Time Evaluation	Jan-May 1971
3. PF and CM Parallel Operation	Oct.-Dec. 1970 & June-Dec. 1971
4. Shock Load Studies	July-Dec. 1971
5. Pilot Plant Operation	June-Oct. 1971

Plant Layout

Construction of the plant proper was completed in April 1970, and air and return sludge piping modifications were finished in July 1970.

The overall plant layout is shown in an aerial view in Figure 2. The activated sludge system consisting of a blower building,



AERIAL VIEW OF PLANT

Figure 2

4 rectangular aeration tanks, and 4 circular final settling tanks is in the upper left hand corner of the picture.

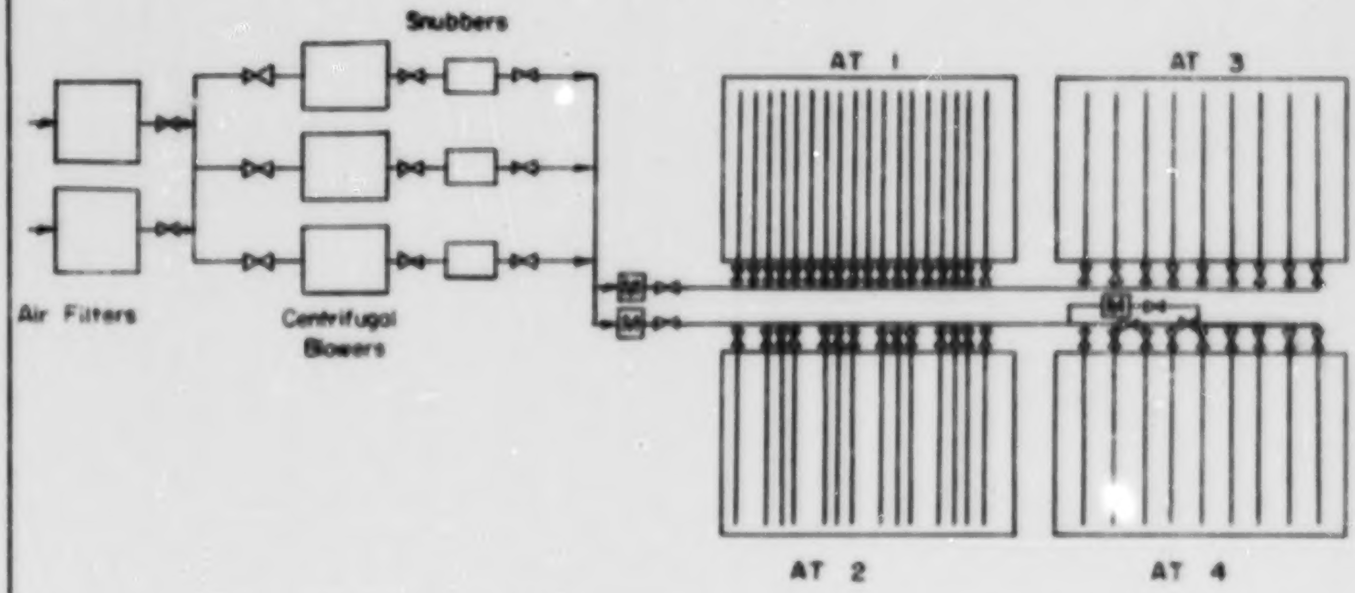
Each aeration tank has dimensions of 57.5' x 25' x 13' water depth (Tank volume excluding effluent channel is 132,000 gal). There are 5 inlet chambers for primary effluent and 4 inlets for return sludge on 1 side of each tank (See Figure 1). Mixed liquor leaves the CM tanks over a longitudinal weir extending the full length of the tanks on the side opposite the inlet chambers. Sluice gates are provided between tanks and on one end to provide for longitudinal flow. The flow configurations shown in Figure 1 are for parallel CM and PF operation. Baffling is shown down the center of aeration tanks 2 & 4. Details of other flow patterns studied are included in the mixing study portion of the report.

The final settling tanks are peripheral inlet-peripheral outlet circular tanks with a diameter of 57' and a water depth of 14' (Tank volume excluding effluent and influent channels is 270,000 gal.). Sludge withdrawal is by means of suction through a moving header at the bottom of the tank.

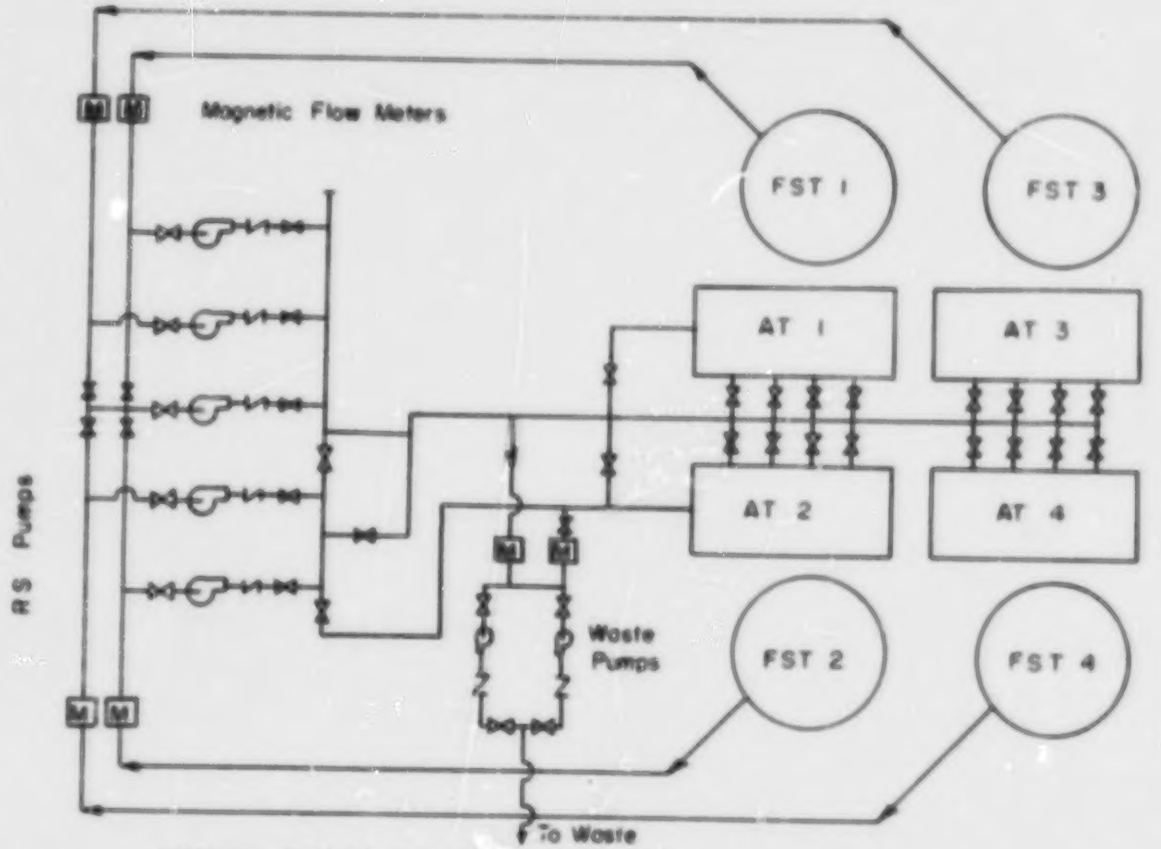
Air is supplied by a centrifugal blower with a maximum capacity of 6,000 cfm. Two identical standby blowers were included in the design. Air is metered separately to each system through differential pressure flow tubes. A diagram of the air piping system is shown in Figure 3. Details of diffuser placement are included in the mixing study phase of the report.

The return sludge system is shown schematically in Figure 4. The valving is such that return sludge from each pair of final tanks can be kept separate. Each waste sludge pump can be valved to pump only from one system. All waste sludge is cycled back to the primary settling tanks and from there is pumped to the anaerobic digestion system. Waste and return sludge flows are measured using magnetic flow meters.

Automatic proportional samples were installed in the flow stream at three points. These points are marked with an X in Figure 1. The unit sampling the primary effluent was placed at the end of a 200' channel to insure adequate time for mixing of the effluent from the two primary clarifiers. Similarly, each final effluent sampler was placed in a manhole at the end of a 150' long, 36" diameter concrete pipe just before entry of the liquid into the chlorine contact system.


Snubbers
AT 1
AT 3
Air Filters
Centrifugal Blowers
AT 2
AT 4

ACTIVATED SLUDGE AIR SYSTEM
Figure 3


Magnetic Flow Meters
M
FST 1
FST 3
AT 1
AT 3
AT 2
AT 4
FST 2
FST 4
RS Pumps
Waste Pumps
To Waste

RETURN SLUDGE SYSTEM – Figure 4

The samplers were of the bucket type and sampled approximately 35 ml at a time. The samplers were actuated on a proportional basis by a signal from a final effluent flow meter tied into a Parshall flume at the outlet of the chlorine contact tank. Tests showed the units to draw a sample approximately every 80,000 gallons of flow.

Project Limitations

The results reported herein, of course, pertain uniquely to the Freeport sewage which is a weak domestic-commercial waste with limited industrial contribution. This can make differences in performance between CM and PF systems since the shock load effect at the head end of the PF aeration tanks was less pronounced than would be the case with stronger wastes.

SECTION IV

AERATION TANK MIXING STUDY

Introduction

Before any parallel comparisons could be made, it was first necessary to establish what combination of flow pattern, diffuser placement, and air application rate would yield a situation approaching completely-mixed on one side versus plug flow on the other side. To accomplish this, a rhodamine dye fluorometer tracer system was developed, tested under laboratory conditions, and used to evaluate the flow regimes in the aeration tanks under varying conditions.

The basic objective of this phase of the study was then to determine the effect on the mixing regime of varying the conditions of:

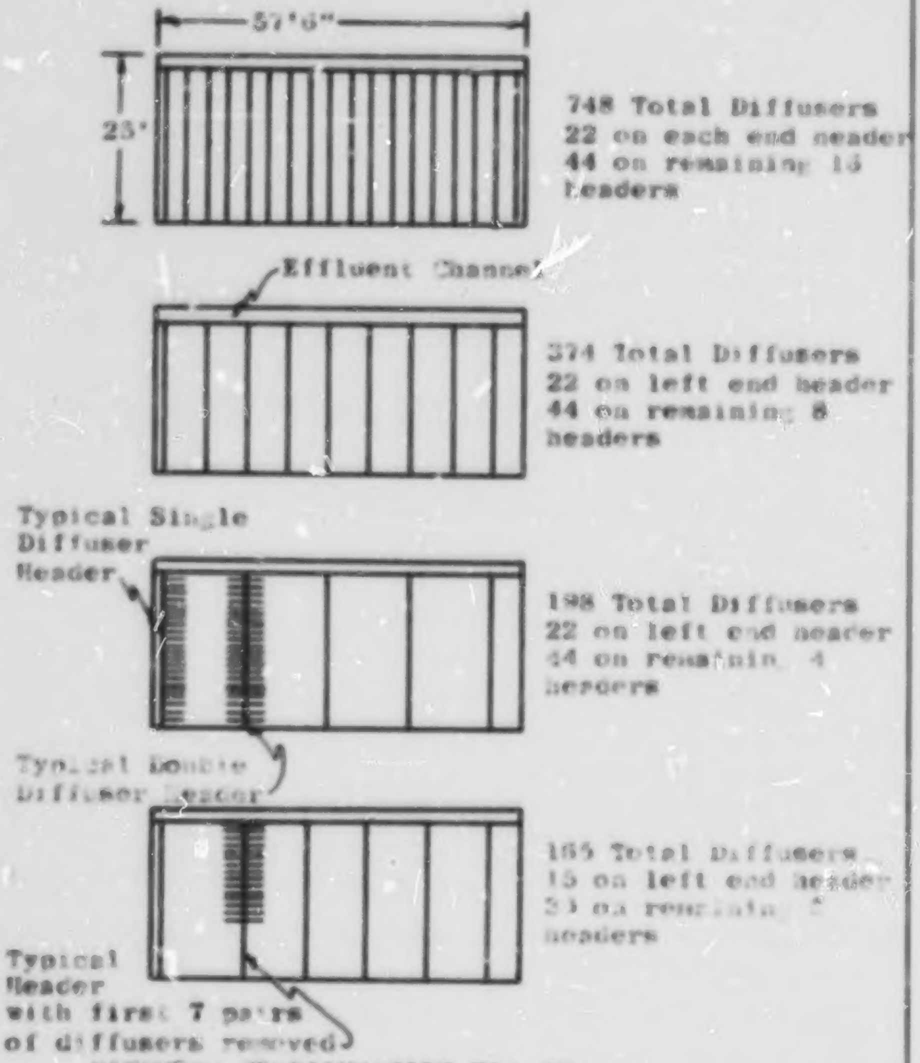
1. Aeration tank configuration(length:width ratio , number of tanks)
2. Influent and effluent arrangement
3. Diffuser placement
4. Influent flow rate
5. Air flow rate

Figure 3 shows the arrangement of air piping to the aeration tanks. Air is metered separately to each side with a venturi-type flow tube utilizing a differential pressure transmitter to an indicating meter on the main control panel. An additional flow tube with an inclined manometer was inserted to measure the air split between aeration tanks 2 & 4. The system of butterfly valves allows for balancing of air and changing of diffuser header configurations without draining the tanks.

The flexibility features utilized in the mixing study can then be summarized as follows:

1. Multiple inlet points for primary effluent and return sludge
2. Side and end effluent weirs
3. Sluice gates between tanks
4. Individually valved air headers with provision for creating different diffuser configurations
5. Variable output blowers and individual metering of air to each pair of tanks

The diffusers were of the nylon sock type with a diameter of 3" and a length of 2'. Normal arrangement on the headers were pairs of diffusers on opposite sides of the header spaced one foot apart. Figure 5 shows the diffuser config-


57'6"
25'
748 Total Diffusers
22 on each end header
44 on remaining 16
headers
Effluent Channel
374 Total Diffusers
22 on left end header
44 on remaining 8
headers
Typical Single
Diffuser
Header
198 Total Diffusers
22 on left end header
44 on remaining 4
headers
Typical Double
Diffuser Header
165 Total Diffusers
15 on left end header
30 on remaining 5
headers
Typical
Header
with first 7 pairs
of diffusers removed

DIFFUSER CONFIGURATION FOR 2M
AERATION TANK MIXING STUDY

Figure 5

urations used for the CM portion of the study. Figure 6 outlines similar information for the PF study.

Experimental Procedure - Tracer System

The fluorometer system is depicted schematically in Figure 7 and in a photograph in Figure 8. The centrifugal pump draws a continuous sample (at a rate of 10gpm) from the desired sample point into the de-aeration chamber. Entrained air escapes in the chamber and any excess sample overflows. The diaphragm pump withdraws a continuous sample from the de-aeration chamber and transfers it to the fluorometer flow cell at a rate of 150 ml/min. The relative intensity of fluorescence of the sample is measured by the fluorometer, and the output is recorded on the strip chart recorder.

The flow system used was found to be necessary to eliminate extreme variability in output due to entrained air passing through the flow cell at a high velocity. This would occur if the centrifugal sample pump were allowed to discharge directly to the flow cell.

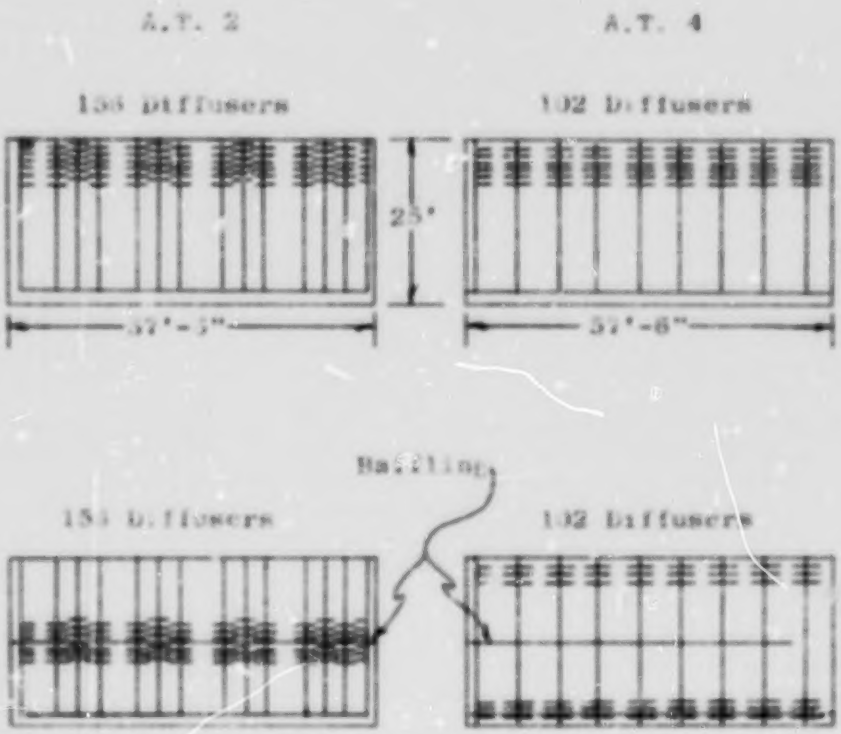
Two methods were used to determine the lag time of the system. First, dye was used as a tracer. A twenty-gallon tank was filled with tap water and dye was added to the water and mixed thoroughly. The intake of the diaphragm pump was placed in the liquid and the pump was started. The time until an initial reading was obtained on the fluorometer was measured. In each case this was 30 seconds.

The additional time until a peak reading was obtained was measured, and this averaged 1.0 minute for 3 runs with dye concentrations of 50, 60 and 70 micrograms per liter.

Second, the volume of the system, i.e., diaphragm pump, connecting hoses and the flow cell was computed and found to be 225 milliliters. At a flow rate of 150 ml/min the theoretical detention time would be 1.5 minutes. If the additional volume of the centrifugal pump, de-aeration chamber and connecting hose is calculated, this is 1.2 gallons. At a flow rate of 10gpm, the theoretical detention time in this part of the system would be 0.12 minutes. This was considered negligible; therefore, the lag time of the system was established at 1.5 minutes.

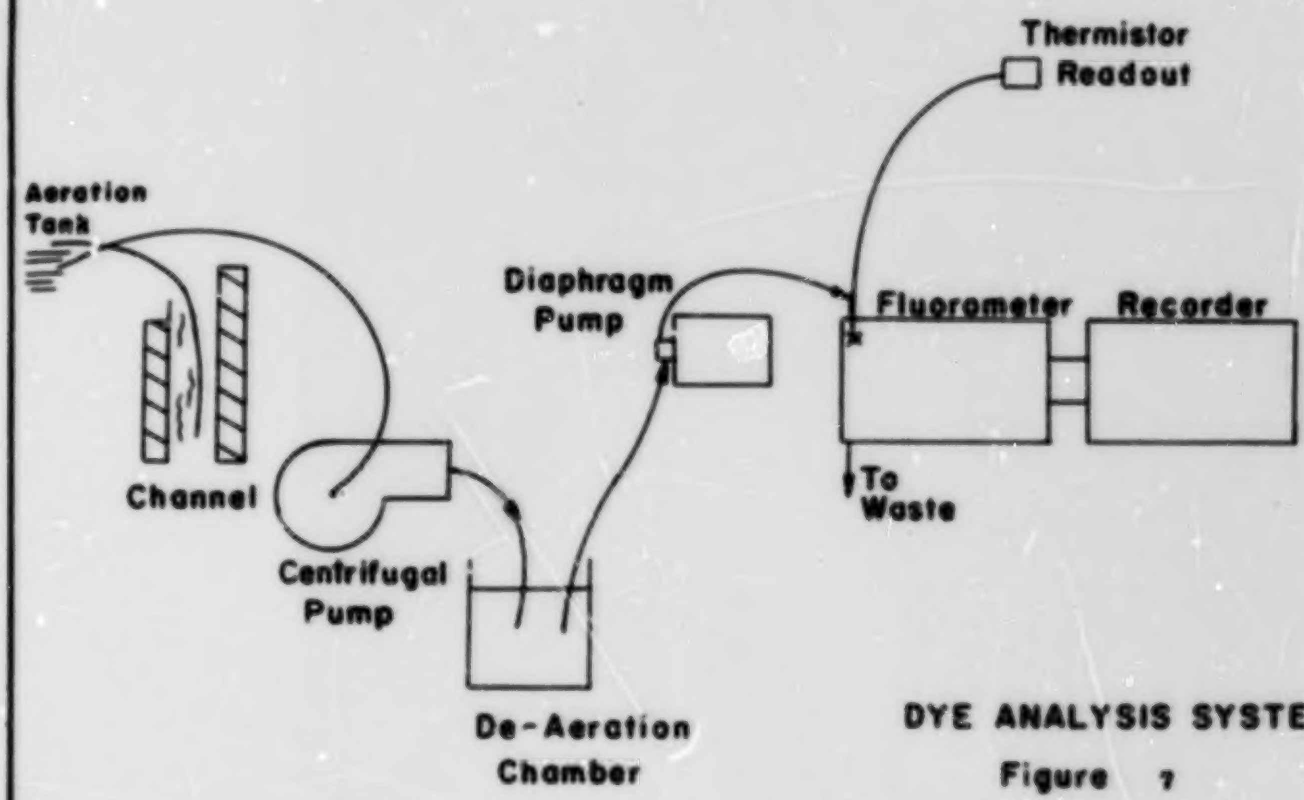
To calibrate the fluorometer system, the arrangement shown in Figure 9 was developed. This arrangement was used to duplicate the field operation of the dye analysis equipment to the point of the de-aeration chamber.

The ice water bath was used to control the liquid tempera-


A.T. 2
A.T. 4
156 Diffusers
102 Diffusers
25'
57'-3"
57'-6"
Baffling
156 Diffusers
102 Diffusers

Diffuser Configurations For FF
AERATION TANK MIXING STUDY

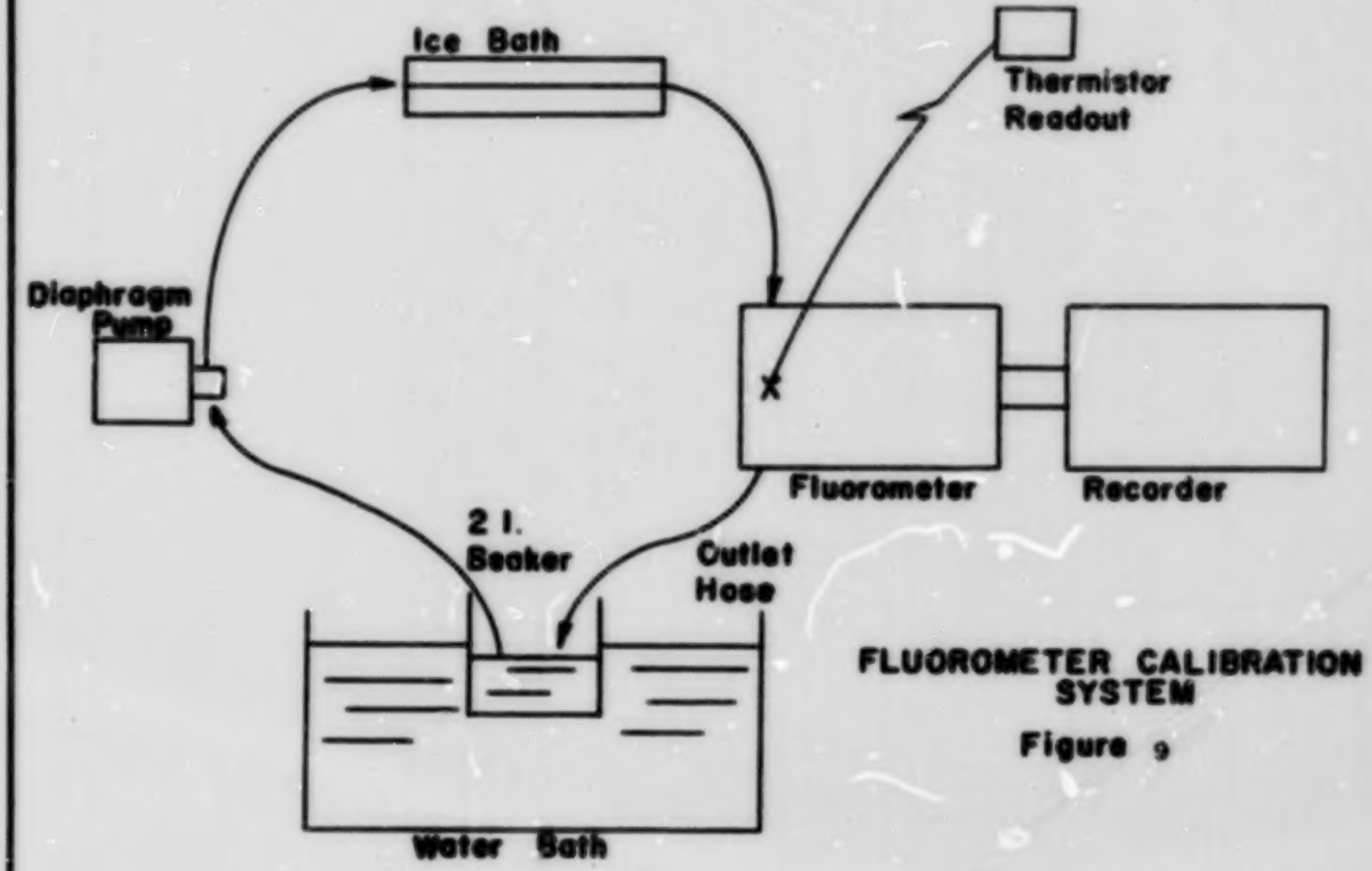
Figure 6


Aeration Tank
Channel
Centrifugal Pump
De-Aeration Chamber
Diaphragm Pump
Thermistor Readout
Fluorometer
Recorder
To Waste

DYE ANALYSIS SYSTEM

Figure 7

DYE ANALYSIS SYSTEM
IN FIELD
FIGURE 8


Ice Bath
Thermistor Readout
Diaphragm Pump
Fluorometer
Recorder
2 l. Beaker
Outlet Hose
Water Bath

FLUOROMETER CALIBRATION SYSTEM

Figure 9

ture entering the fluorometer flow cell. This temperature was measured within the cell itself by inserting a thermistor with connecting leads in the outlet hose, so the thermistor tip stopped inside the cell just below the glass viewing window.

The procedure for a typical calibration test is outlined below:

1. Standard solutions of rhodamine B or WT dye are prepared. The standard solution concentration normally used was 2 mg/l.
2. The hoses, pumps and flow cell are flushed with test liquid containing no dye. The system is drained. A volume of test liquid is added to the solution beaker. This volume is such that when standard rhodamine dye solution is added, the total volume in the system will be two liters.
3. The pump is started and liquid is circulated through the system. The ice bath is used to adjust the temperature. When the desired temperature is reached, the fluorometer is zeroed.
4. The desired quantity of dye solution is then added to the solution beaker using a volumetric pipet.
5. When a constant reading is obtained on the fluorometer at the desired temperature, this relative intensity reading is used to establish a calibration point for the dye concentration in the test solution.
6. If points at other temperatures are desired, the ice bath is used to change the temperature and additional relative intensity readings are taken.

A series of runs were made to determine the effect of temperature on fluorescence. Figure 10 shows the curve used for temperature correction in this study. A similar curve as reported by the U.S. Geological Survey[1] and Butts[2] of the Illinois State Water Survey is shown. This latter curve is developed from the following relationship presented by Butts:

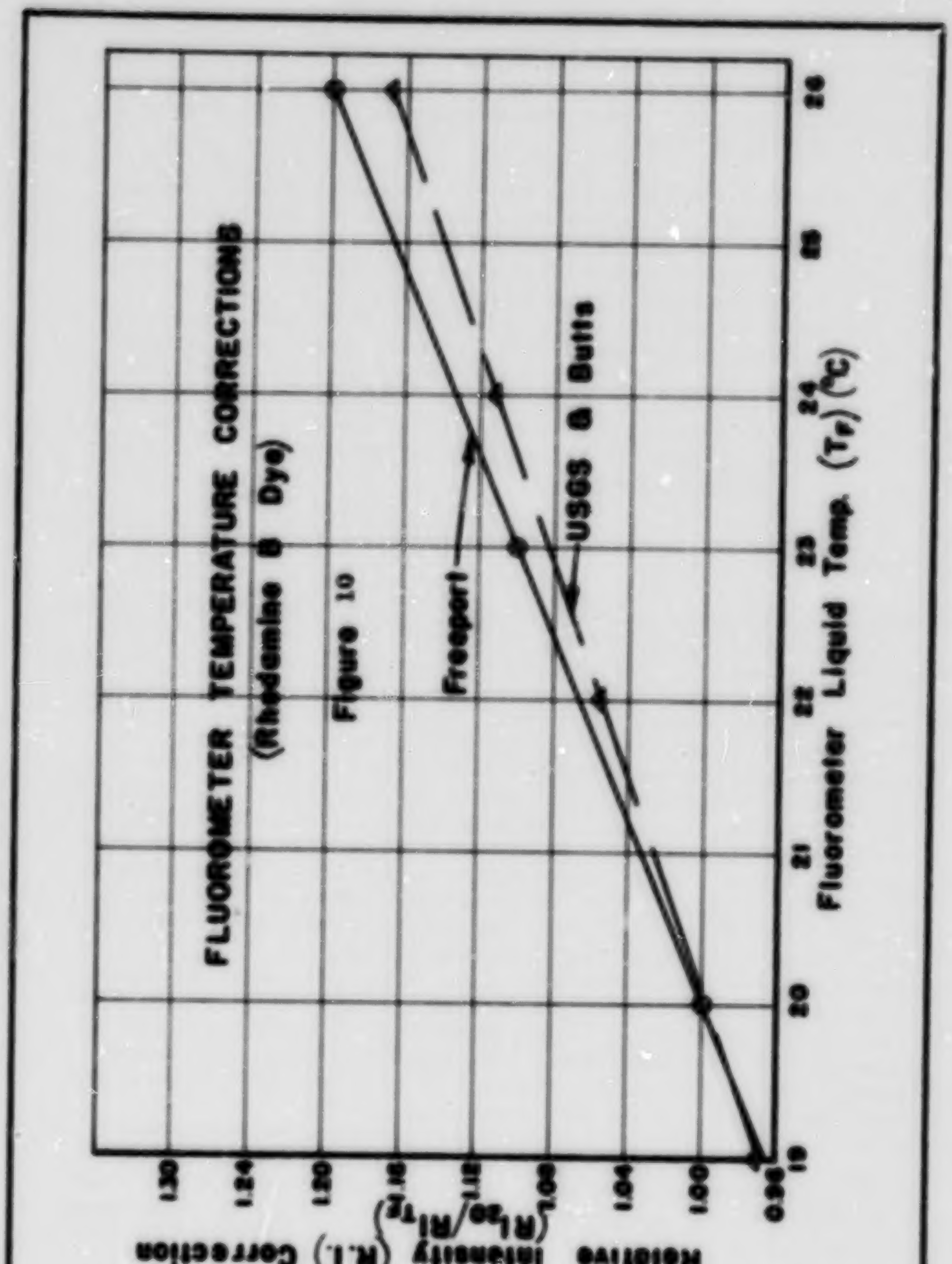
$$F_t = F_s \, e^{n(T_s - T_t)}$$

where F_t = fluorescence at test temp , T_t (°C)

F_s = fluorescence at standard temp , T_s (°C)

$n = 0.027 \ ^{\circ}C^{-1}$

It should be noted that a Turner fluorometer was used in the experimental temperature studies by Butts while an



FLUOROMETER TEMPERATURE CORRECTIONS

(Rhodamine B Dye)

Figure 10

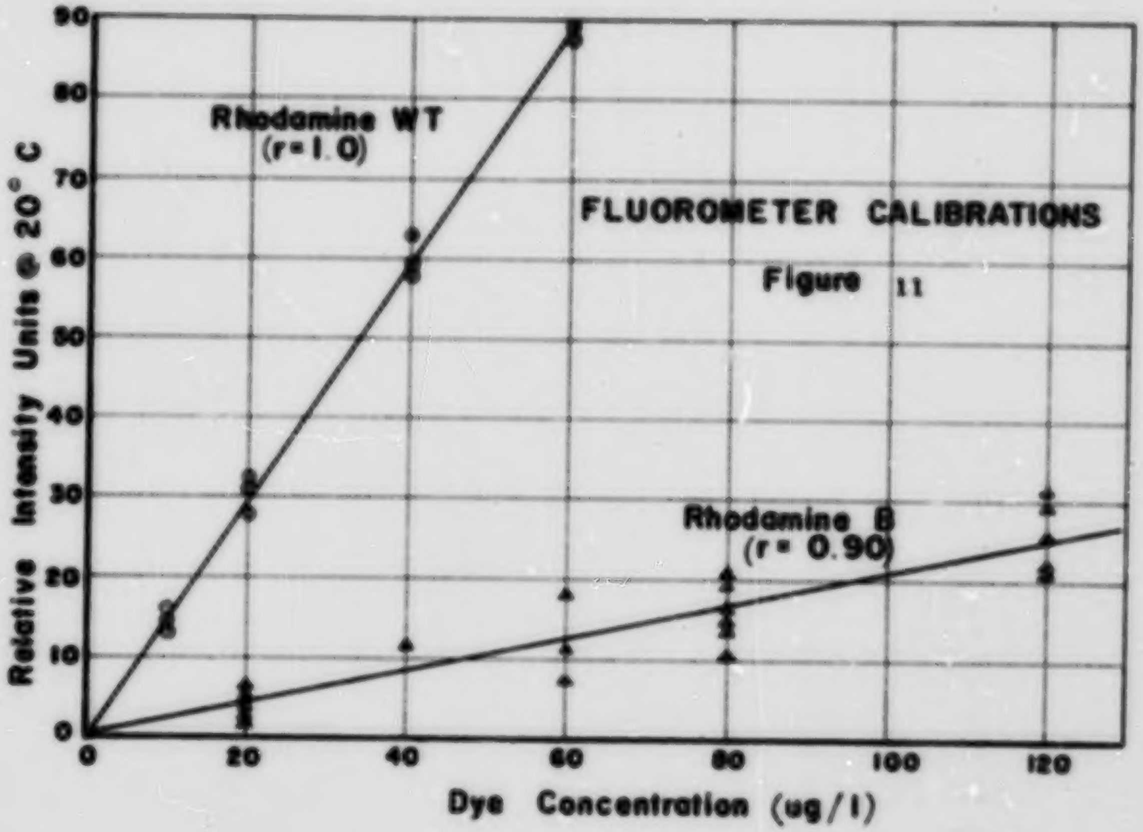
Aminco fluorometer was used in this study.

Laboratory testing of the Freeport system indicated that the calibration curve was linear in the range of 0-100 micrograms/liter, but that the slope varied for each run using rhodamine B dye.

To evaluate the variability in calibration during each field tracer test, samples of primary effluent were taken periodically as it entered the aeration test tank. These samples were composited and used as the test liquid for a laboratory calibration test as described above. Figure 11 shows the results of this calibration evaluation for rhodamine B and WT dye. The lines shown were drawn using a least squares technique and the respective correlation co efficients (r) show that rhodamine WT demonstrates less variability in calibration. This is probably due to its characteristic of less adsorption on suspended solids.

For each field test, the experimental procedure was as follows:

1. The dye analysis system (Figure 7) was positioned near the effluent weir of the test aeration tank, and the suction hose of the sample pump was placed just inside the effluent weir at a water depth of approximately 1.5'.
2. Prior to the start of a run, the air and primary effluent flow rates were adjusted to constant levels. The air flow was controlled with a butterfly valve on the blower suction and measured through a venturitype flow tube in the air line. Liquid flow was held constant either by setting a variable speed raw sewage pump or by adjusting stop gates into other tanks to by-pass excess flow. The liquid flow rate was measured via a Parshall flume located at the effluent of a final settling tank in series with the test aeration tank.
3. In each case, dye was added as a pulse input at each liquid inlet point by releasing it by hand from a one gallon glass bottle. Zero time was recorded as the moment the dye was added. In computations, the true zero time was shifted ahead 1.5 minutes to account for the lag time of the dye analysis system.
4. During the run, air and sewage flows were recorded every 15-20 minutes, and the temperature of liquid in the fluorometer flow cell was checked every 5 minutes. Samples of the primary effluent entering the tank were collected periodically to be used for calibration purposes.



FLUOROMETER CALIBRATIONS

Figure 11

Three basic aeration tank configurations were tested. They are outlined in Figure 12 as:

I. Complete-mix consisting of transverse flow with single or multiple inlet points, all-over-the-bottom diffuser placement, and a side effluent weir (length:width = 0.43:1).
II. Plug-flow consisting of longitudinal flow through two tanks in series with one pair of end inlet points, diffusers placed along one side and an end effluent weir (length:width = 4.6:1).
III. Plug-flow with temporary baffling changing the length:width ratio to 18.4:1, side diffuser placement, one pair of inlet points and an end effluent weir. Flow pattern III was created by installing a frame-work covered with neoprene coated nylon in the center of the same tanks used for pattern II (see Figure 13).

Types of Analysis Used

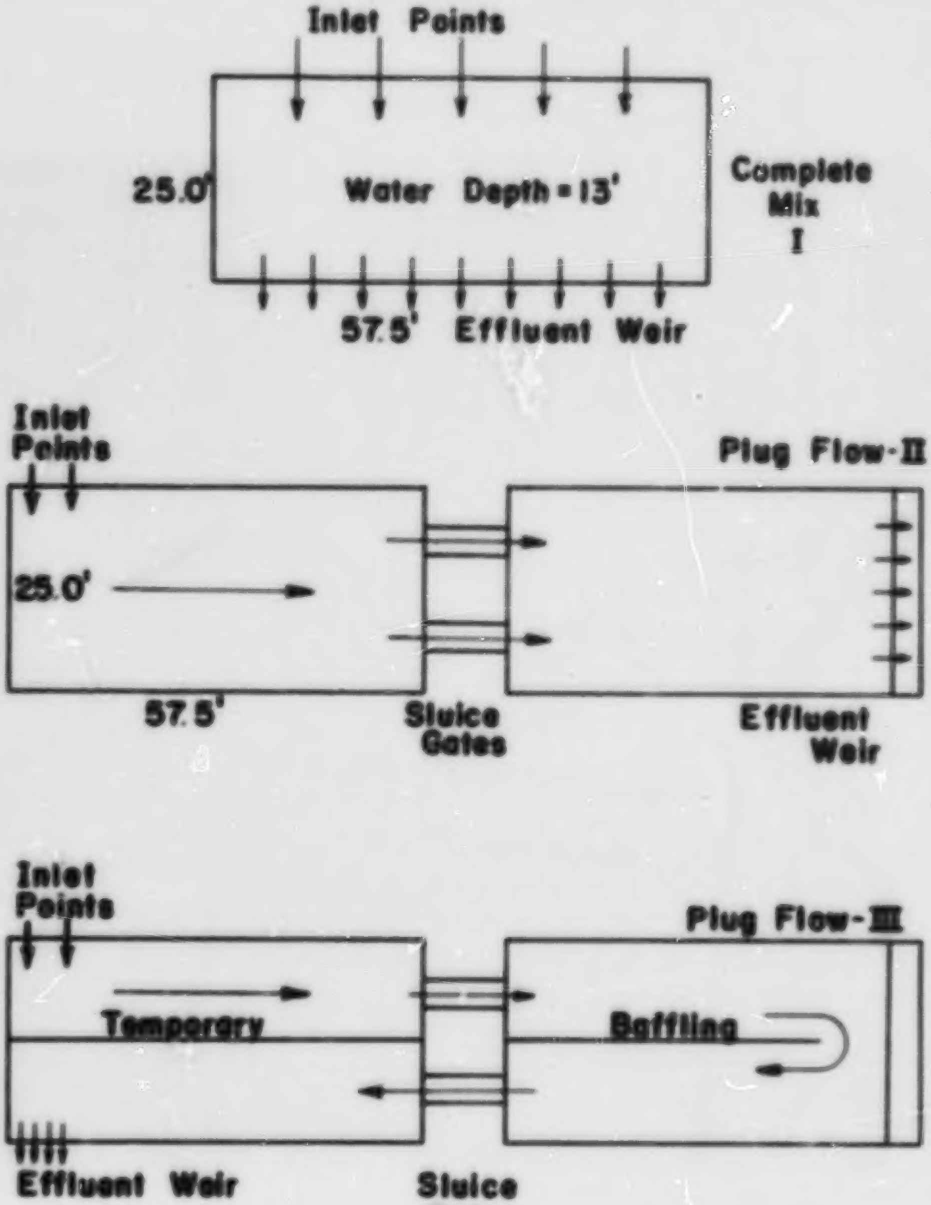
The primary approach used to analyze the data is outlined by Levenspiel[3]. If a pulse input of dye is added to the influent stream of a closed system and the concentration of dye in the effluent stream is measured with respect to time, a curve can be plotted which relates these two parameters. The curve is called a "C" curve and is defined as follows:

t - time from moment of dye addition
C - dye concentration at time t
C_0- initial concentration of dye if it is evenly distributed throughout the vessel
T - mean residence time of liquid in vessel
θ - t/T
$f(\theta)$ - C/C_0

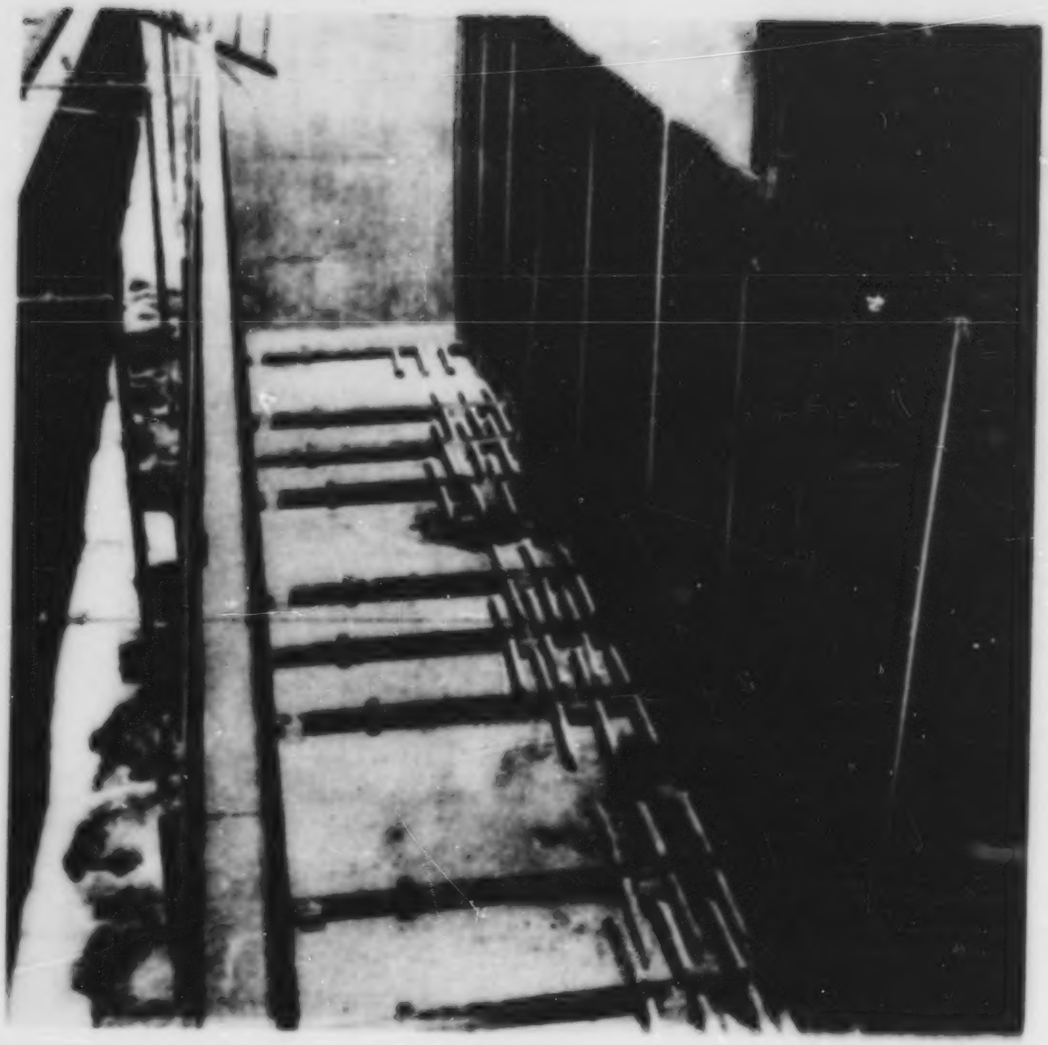
From a plot of θ versus $f(\theta)$, it is possible to calculate the mean, μ, and the varience, σ^2. The definitions for a discrete number of points are as follows:

$$\mu = \frac{\sum_{i=1}^{n} \theta_i f(\theta_i)}{\sum_{i=1}^{n} f(\theta_i)} \qquad \theta^2 = \frac{\sum_{i=1}^{n} \theta_i^2 f(\theta_i)}{\sum_{i=1}^{n} f(\theta_i)} - \mu^2$$

If a tanks-in-series model is used to describe the mixing conditions, n values (n refers theoretically to the number of completely-mixed tanks in series in the system) can be computed from σ^2, where $n = \frac{1}{\sigma^2}$.


Inlet Points
25.0'
Water Depth = 13'
Complete Mix I
57.5' Effluent Weir
Inlet Points
Plug Flow-II
25.0'
57.5'
Sluice Gates
Effluent Weir
Inlet Points
Plug Flow-III
Temporary
Baffling
Effluent Weir
Sluice Gates

AERATION TANK FLOW PATTERNS
Figure 12



TEMPORARY BAFFLING
IN AERATION TANK 2

FIGURE 13

An n value 1 indicates ideal complete mixing and an n value of ∞ represents ideal plug flow.

However, Timpany[4] states that to accurately determine σ^2, a test must be extended to at least ten detention times. This is impractical due to the length of time involved in full scale testing.

Fair, Geyer, and Okun[5] define n in the following manner:

$$\frac{n-1}{n} = \theta_{max}, \text{ where } \theta_{max} = \theta \text{ at maximum } f(\theta) \text{ for a unimodel curve}$$

This was the approach used to calculate n.

The following steps were incorporated into a computer program and used to compute the mean residence time in the vessel n, and per cent dye recovery.

1. Compute mean residence time -

$$\bar{T} = \frac{\sum_{i=1}^{n} t_i C_i}{\sum_{i=1}^{n} C_i}$$

2. From $\bar{T}$ compute θ_i -

$$\theta_i = \frac{t_i}{\bar{T}}$$

3. From $\bar{T}$ compute effective tank volume, $\bar{V}$ -

$$\bar{V} = Q_{Avg}\,\bar{T}$$

4. Compute C_o -

$$C_o = \frac{\text{Dye weight added}}{\bar{V}}$$

5. Compute $C_i/C_o = f(\theta)$ for equal intervals of θ

6. Find θ max

7. $n = \dfrac{1}{1 - \theta_{max}}$

8. Compute per cent dye recovery -

$$\text{weight of dye recovered} = \sum_{i=1}^{n} C_i \Delta t \times Q_{avg}$$

$$\%\ \text{recovery} = \frac{\text{weight recovered} \times 100\%}{\text{weight added}}$$

Theta $(\Theta)_{max}$ can also be used to determine the dispersion parameter, $\frac{D}{uL}$, used by Murphy and Boyko[6].

Levenspiel[3] defined $\frac{D}{uL}$ as outlined below:

$\frac{dc}{dt} = D \frac{d^2c}{dx^2}$, defines D where c = concentration at a point x and a time t in a vessel

L = vessel length
u = mean displacement velocity along vessel length

For complete mixing conditions $\frac{D}{uL}$ approaches infinity, and for plug flow conditions $\frac{D}{uL}$ approaches zero. Timmany[4] presents curves relating $\frac{D}{uL}$ and Θ max. Levenspiel[3] relates the parameters n and $\frac{D}{uL}$ by the following equation:

$$1/n = \sigma^2 = 2\frac{D}{uL} - 2\left(\frac{D}{uL}\right)^2 \left(1 - e^{-uL/D}\right)$$

The second approach in evaluating the tracer data is to observe the slope of the washout curve. Smith[7] gives the following relationship for an ideal complete mixing tank:

$\frac{dC}{dt} = \frac{-QC}{V}$, where C = concentration in effluent
t = time
V = tank volume
Q = influent flow rate

Integration of this equation yields the following:

$C/C_o = e^{-t/T} = e^{-kt}$, where for completely-mixed systems $k = 1/T$

Since all test data is presented in curves relating C/C_o versus t/T, the following mathematical manipulation was used to evaluate k directly from the curve:

$$k = \frac{\Delta \ln C/C_0}{(\Delta t/T)\ T}$$

Thirdly, a comparison of the configurations can be made utilizing a measure of the extent of dispersion. The Morril Index is defined by Kothandaraman[8] as t_{90}/t_{10}, where

t_{10} = time for 10% of tracer to pass sample point

t_{90} = time for 90% of tracer to pass sampling point

Experimental Results

Table 1 summarizes the data from ten tests made during the summers of 1970 and 1971 using aeration tank configuration I as depicted in Figure 12. Table 2 shows similar data taken during the summer of 1970 for configuration II. Table 3 presents the data from tracer tests for configuration III conducted during the summer of 1971.

The first phase of the study consisting of runs 1-4 and 7 evaluated mixing with 5 inlet points and a transverse flow pattern. Air flow, primary effluent and the number of diffusers were varied.

The values of n for these runs are very close to 1 indicating complete mixing. The time-to-peak approach was used to evaluate n. Figure 14 shows the relationship between n and sewage flow rate for the multiple inlet configuration. Figure 15 shows a similar relationship between n and air flow rate for the same runs. The plots indicate a slightly increasing n (deviation from ideal complete-mixing) for increasing air and liquid flow rates. However, it should be noted that for each run both flow rates increased or decreased simultaneously. It would appear from the trend of the data, that increasing liquid flows tended to over - ride increased back-mixing due to air flows and cause the system to deviate from complete-mixing. A reduction in the number of diffusers (run 7) did not appreciably change the n value.

Runs 5,6,14,15 and 16 were also made using only 1 inlet chamber. Again, the time to peak dye concentration was very short yielding a low Θ max and correspondingly low values of n and $\frac{D}{uL}$.

There are factors other than n which must be considered to completely analyze the mixing pattern. One such factor is k. Table 1 lists k values computed from the slope of the tracer

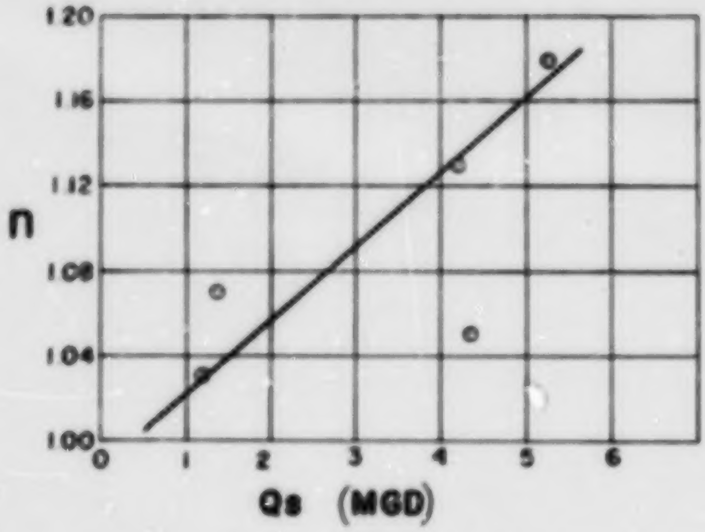
Run	Qs mgd	Qa scfm	No. of Dif-fusers	Inlet Points	n	$\frac{D}{UL}$	K (min^{-1})		Det Time (min)		% DYE Recovery
							Meas.	Theoret.	Meas.	Theoret.	
1	4.35	3700	748	5	1.05	10.5	.0262	0.0216	38	46	186
2	1.37	1110	748	5	1.07	7.0	.0098	0.0068	102	147	95
3	5.26	6940	748	5	1.18	2.5	—	—	—	38	93
4	4.20	3740	748	5	1.13	3.5	.0268	0.0208	37	48	102
5	1.06	925	748	1, End	1.02	31.0	—	—	—	190	90
6	1.14	730	748	1, Cntr	1.04	16.0	—	—	—	172	77
7	1.21	1100	374	5	1.03	24.0	.0119	0.0060	84	167	53
14	3.04	710	216	1, Cntr.	1.01	75.0	.0144	0.0160	69	62	82
15	2.97	1130	408	1, Cntr.	1.00	∞	.0156	0.0156	64	64	80
16	0.99	710	216	1, Cntr.	1.04	26.0	.0088	0.0052	114	192	68

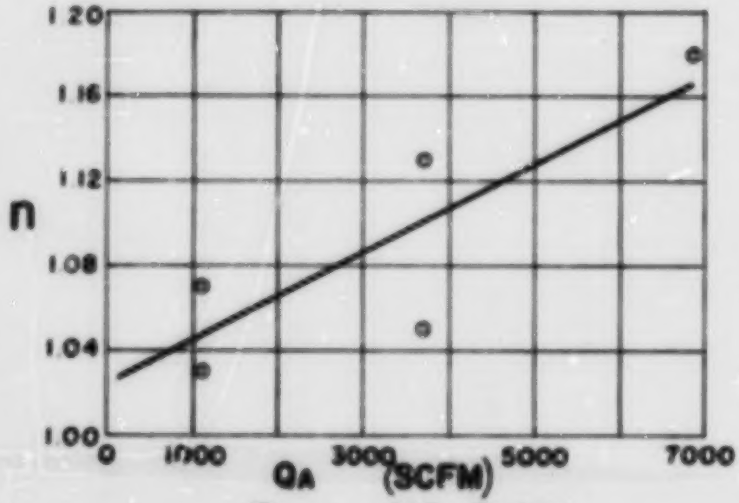
MIXING STUDY PARAMETERS
CONFIGURATION I - COMPLETE-MIX
Table I

MIXING STUDY PARAMETERS
PLUG FLOW

Run	Qs mgd	QA scfm	No. of Diffusers	Inlet Points	n	$\frac{D}{uL}$	Mean Residence Time (min)	Theoretical Detention Time (min)	% Dye Recovery	
8	1.97	1980	156+102	2, End	3.50	0.14	121	193	76	CONFIG.
9	1.02	1110	156+102	2, End	2.98	0.16	145	372	42	II TABLE 2
10	1.25	750	102	1, End	1.21	2.0	89	152	136	
11	3.08	830	156+102	2, End	—	—	112	124	84	CONFIG.
12	3.02	1000	156+102	2, End	10.3	0.056	133	126	106	III TABLE 3
13	1.99	930	156+102	2, End	18.9	0.046	214	191	100	

n vs. SEWAGE FLOW
Figure 14





n vs. AIR FLOW
Figure 15

washout curve. In every case but two (14 & 16), the measured k value was higher than the theoretical k. This indicates that the pulse input of tracer was washing out faster than expected if the entire tank volume were being utilized. The fact that low values of n were obtained and the slope of the washout curve was a straight line on a semilog plot indicated that some tank volume was completely-mixed. However, the higher measured k indicates that this volume is less than the actual volume.

An example of the discrepancy in k is shown for run 2 in Figure 16. The plot was made assuming C_0 equal to the minimum C obtained and T equal to the theoretical detention time. All the data were initially analyzed using these assumptions.

A computer program procedure was then developed from Levenspiel[3] to evaluate the mean residence time and effective volume. The initial concentration, C_0 is then the concentration of dye obtained if the total weight added is distributed throughout this effective volume and T is the mean residence time. Figure 17 shows the same C curve for run 2 from the computer data. Theta (Θ) is expanded because the mean residence time was calculated to be less than the theoretical (k is the same for each method).

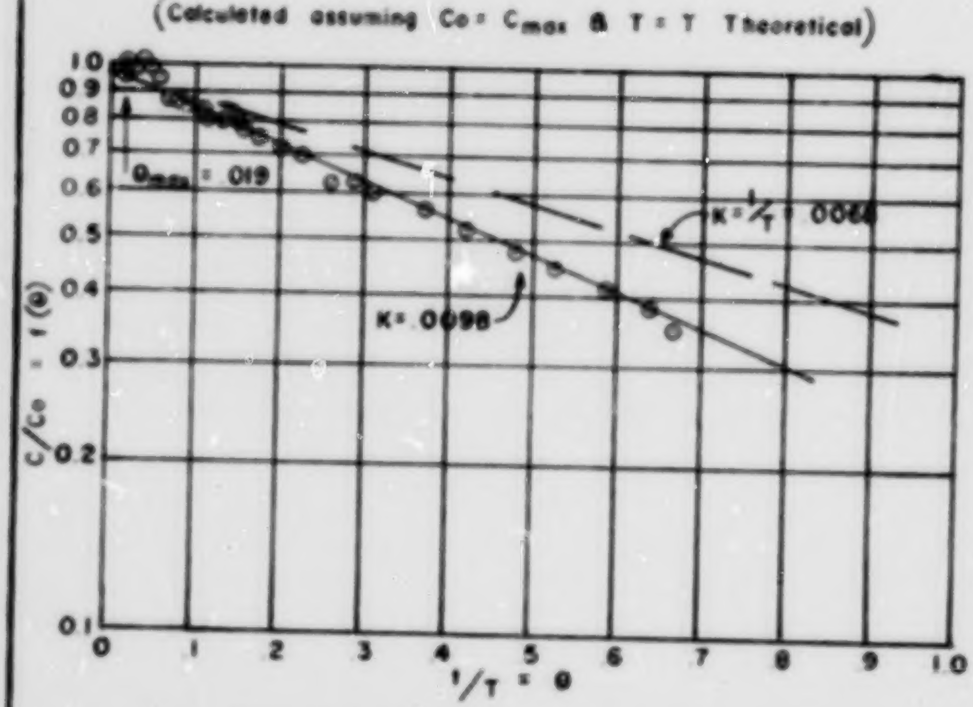
If the mean residence time is computed from the formula,

$$\bar{T} = \frac{\sum_{i=1}^{n} C_i \; t_i}{\sum_{i=1}^{n} C_i}$$

and the run is not extended out several detention times, the data curve is truncated and $\bar{T}$ is calculated to be lower than it actually is (this is the same problem encountered in using variance to calculate n).

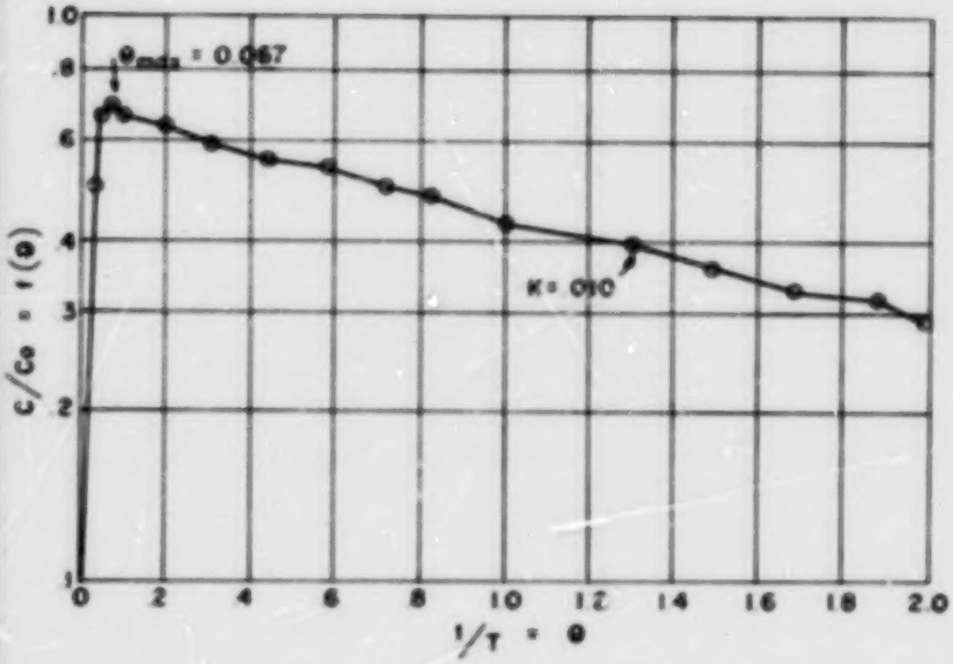
An example of this for run 2 is shown in Figure 18. The data points stop at 120 minutes. The curve was extended along the dotted line by calculating C values from the k established in Figure 16. Applying the mean residence time formula to the extended curve results in a $\bar{T}$ of 89 minutes. If the curve is extended to 500 minutes, $\bar{T}$ calculates to be 104 minutes.

An easier method of establishing $\bar{T}$ for a complete-mixing system is to use the definition of k ($k = 1/T$ or $T = 1/k$). For run 2, $k = .0098$ and $t = 102$ minutes. Using this



C - CURVE - RUN 2
CONFIGURATION 1 - COMPLETE MIX
Figure 16

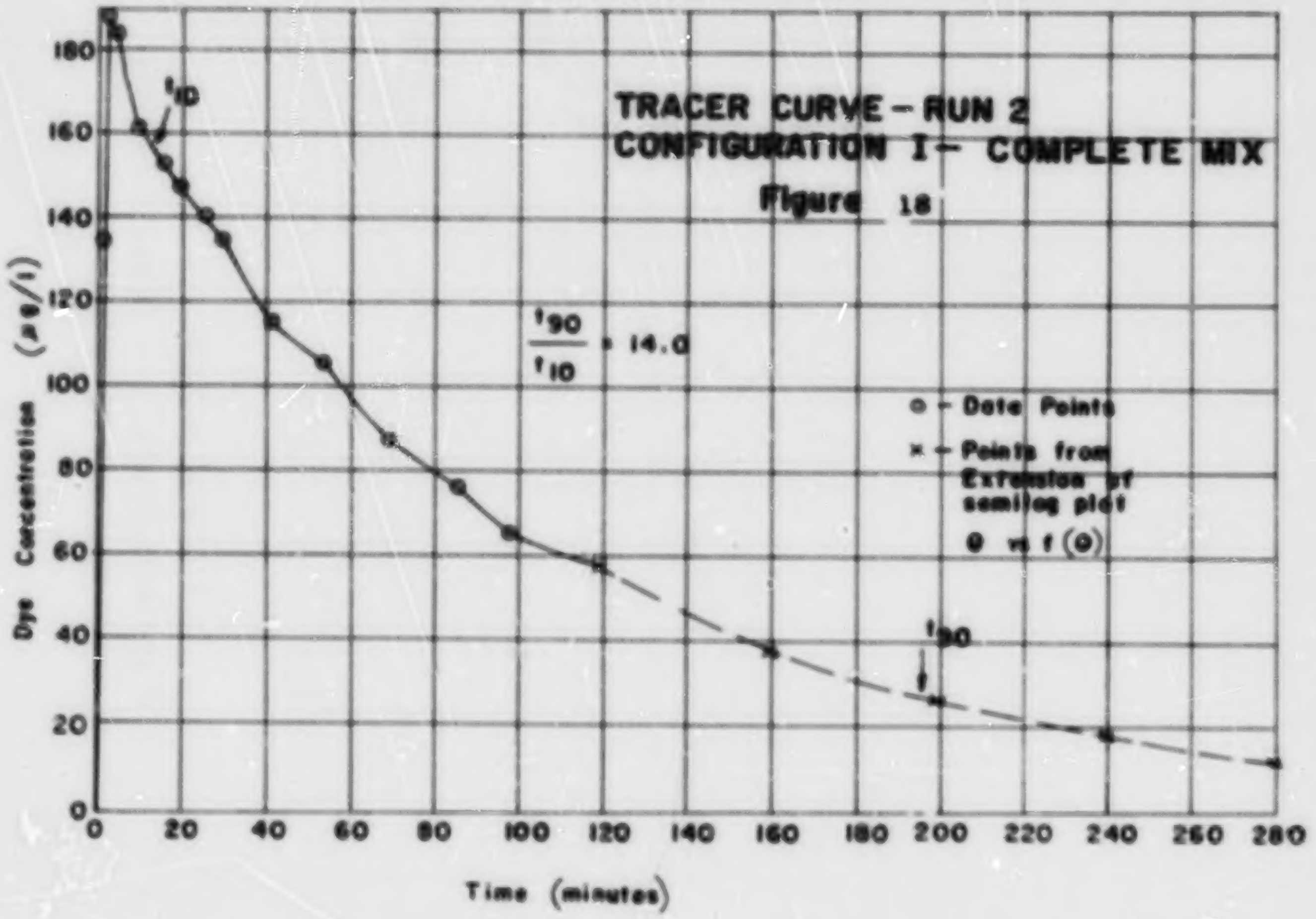
(Calculated from Computer Program)



C – CURVE – RUN 2

CONFIGURATION I – COMPLETE MIX

Figure 17


TRACER CURVE - RUN 2
CONFIGURATION I - COMPLETE MIX
Figure 18

method, the measured detention times are recorded in Table 1. Values of k are not recorded for runs 3, 5,and 6 because the decay curve exhibited 1 or 2 changes in slope.

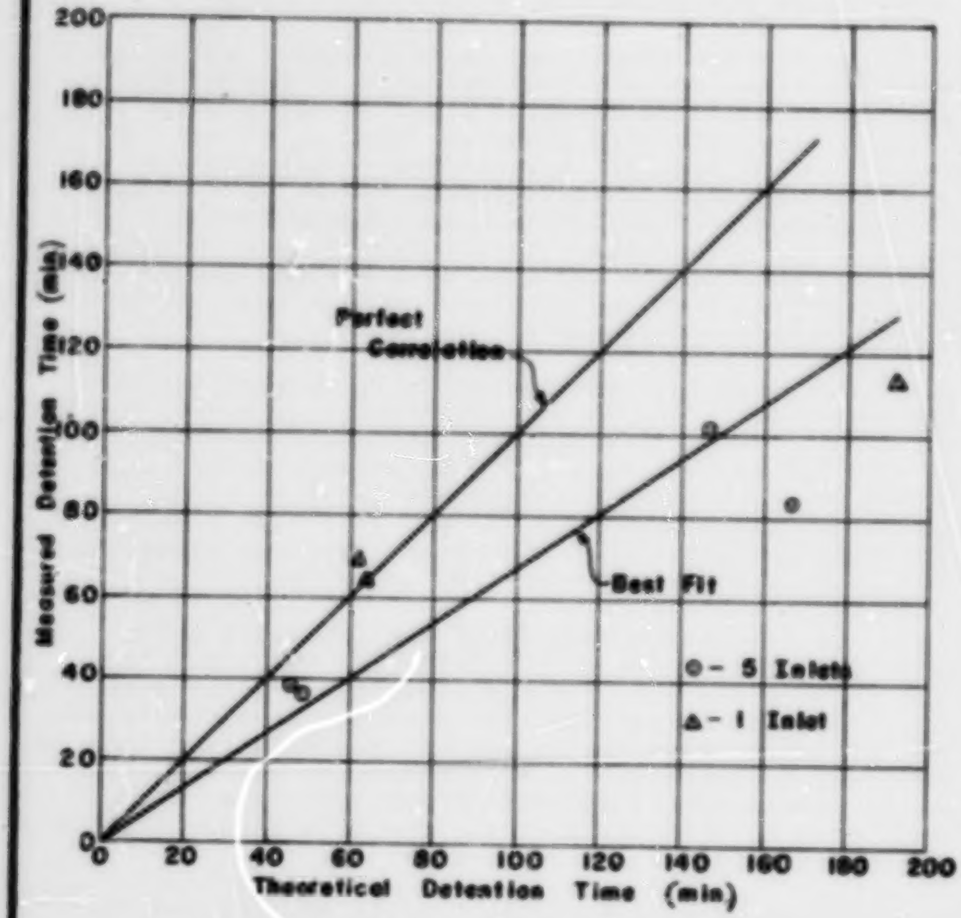
Figure 19 presents a comparison of theoretical and measured detention times for all runs except 3,5, and 6. Scatter in data points would be expected since conditions other than liquid flow rate were changing from run to run. By drawing a line of best fit through the points and forcing it to go through zero, a relationship can be established. This line indicates that for an given theoretical detention time (V/Q), the actual time will be 2/3 of this. A possible explanation would be the creation of a smaller completely-mixed zone riding over the aeration pattern result from the all-over-the-bottom diffuser placement.

It was anticipated that configuration II, a longitudinal flow pattern with a length-to-width ratio of 4.6:1, would yield a typical plug flow pattern. Diffusers were spaced along one side for a 6 foot width to provide spiral flow aeration. However, Table 2 indicates that with normal ranges of air and sewage flow, n values of 1 to 3 were obtained. This indicates that the actual physical situation was one of 2 completely-mixed tanks in series.

Figure 20 is an example (run 8) of the type of C curve obtained from configuration II. Theta (Θ) max was 0.715 and the n value was 3.5. Since a logarithmic decay does not occur initially when the flow deviates from complete-mixing, it is not possible to compute the detention time on the basis of k. Instead, it is necessary to use the mean residence time approach and extend the test data until C approaches zero. This was done for run 8 in Figure 21. The calculation yields a $\bar{T}$ of 121 minutes.

Since the prime objective of the overall project was to compare complete-mixing with plug-flow, it was decided to modify configuration II by adding baffling to obtain configuration III shown in Figure 12. This created a longitudinal flow pattern with a maximum length-to-width ratio of 18.4:1. Table 3 shows that this modification was very successful, since n values of 10.3 and 18.9 were obtained. Mean residence time calculations show good agreement between measured and calculated values. Run 11 exhibited a bimodal curve. This was caused by some leakage through the baffle as initially constructed. Therefore values for n or D are not recorded.

Figure 22 shows the C curve for run 12. Theta (Θ) max has a value of 0.903 yielding an n of 10.3. The peak concentration coincided almost exactly with the mean residence

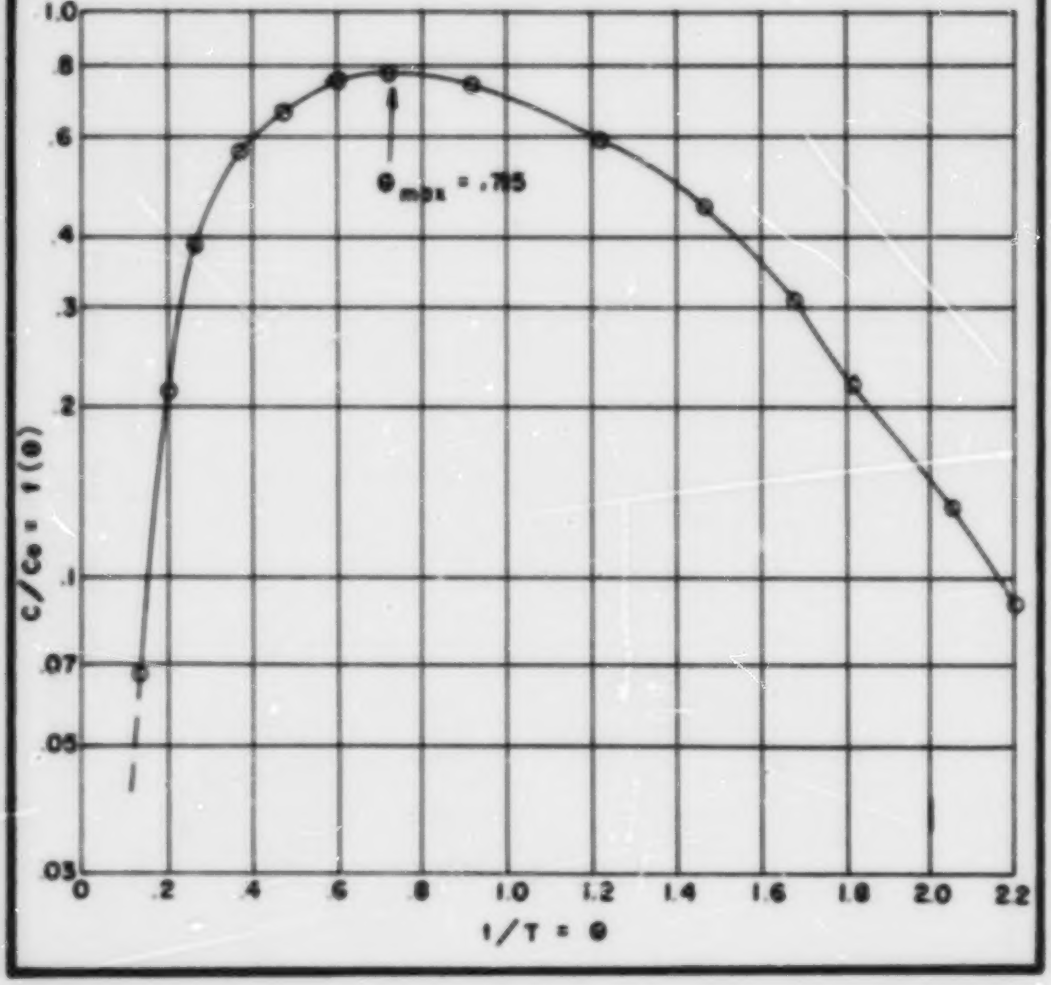


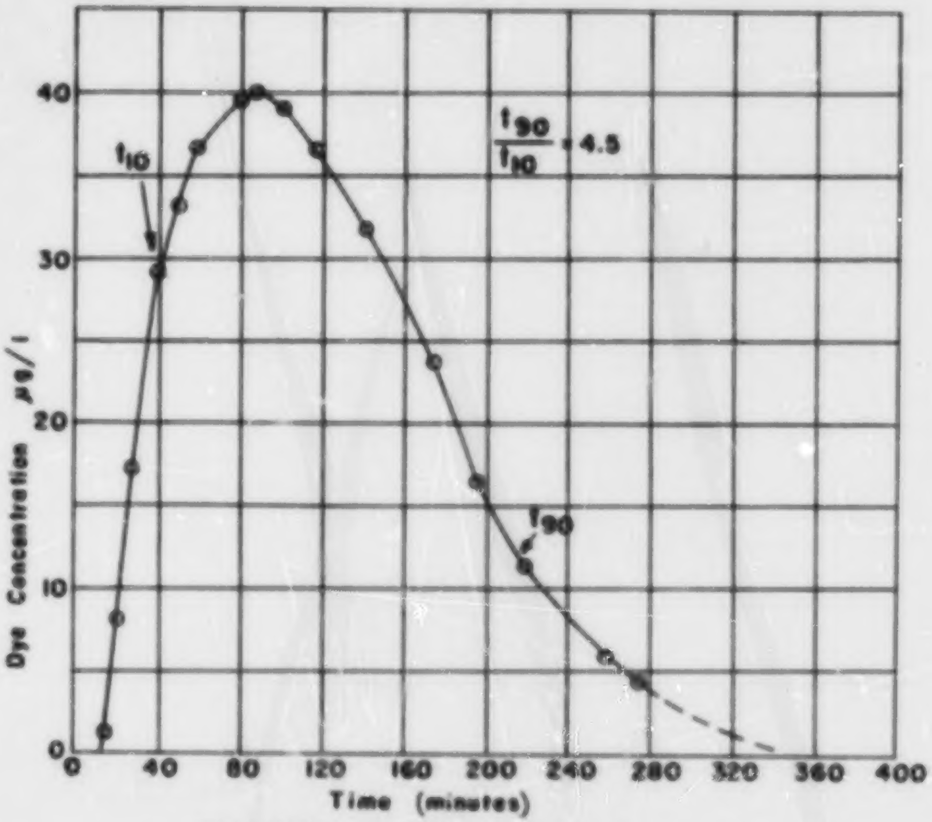
CORRELATION OF THEORETICAL & MEASURED DETENTION TIME COMPLETE MIX

Figure 19

C-CURVE - RUN 8
CONFIGURATION II - PLUG FLOW

Figure 20

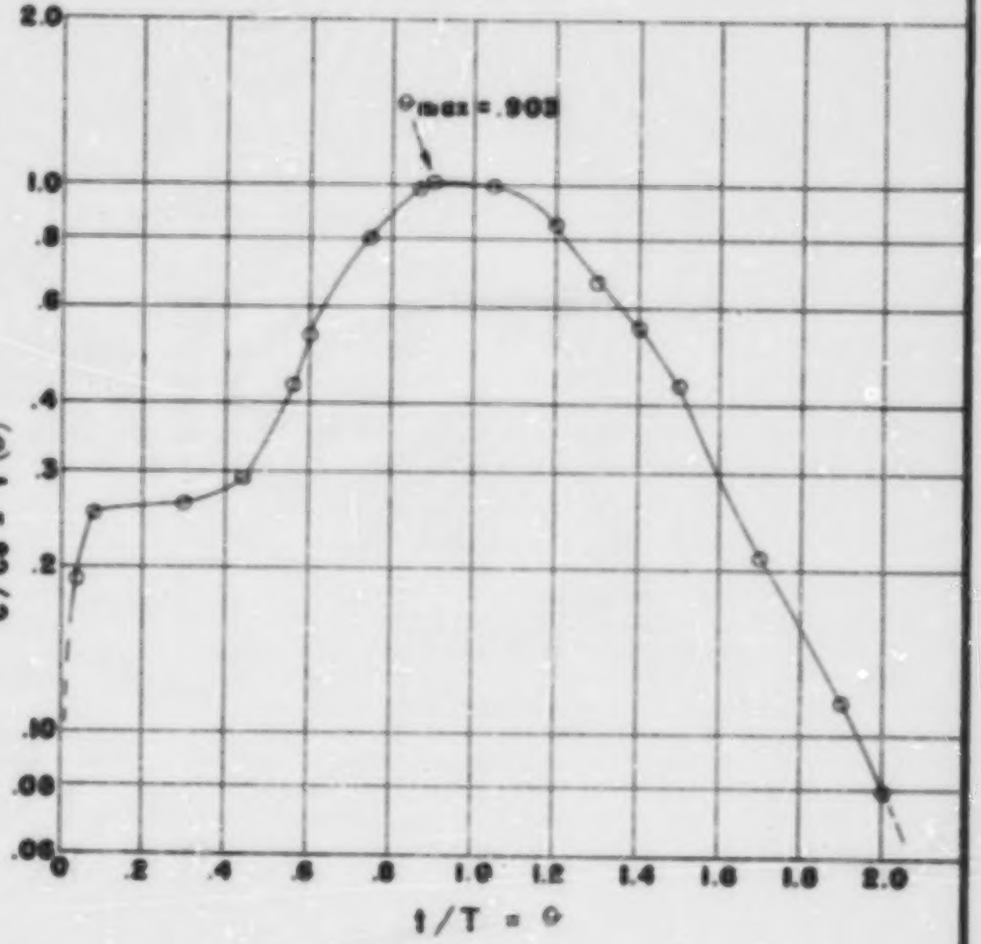




TRACER CURVE – RUN 8
CONFIGURATION II – PLUG FLOW
Figure 21

C-CURVE – RUN 12
CONFIGURATION III – PLUG FLOW

Figure 22



time which also agrees quite well with the theoretical detention time. Configuration III provides a plug-flow pattern for process performance comparison with the completely-mixed pattern of configuration I.

The extent of the longitudinal dispersion for each case can be characterized by the Morril Index, t_{90}/t_{10}. The values are shown for each configuration in Figures 10, 13 and 23. The results are summarized as follows:

Run	Configuration	Morril Index
2	I	14.0
8	II	4.5
12	III	3.8

As would be expected, the greatest dispersion occurred in run 2 with an index of 14.0 and the least occurred in run 12 with an index of 3.8.

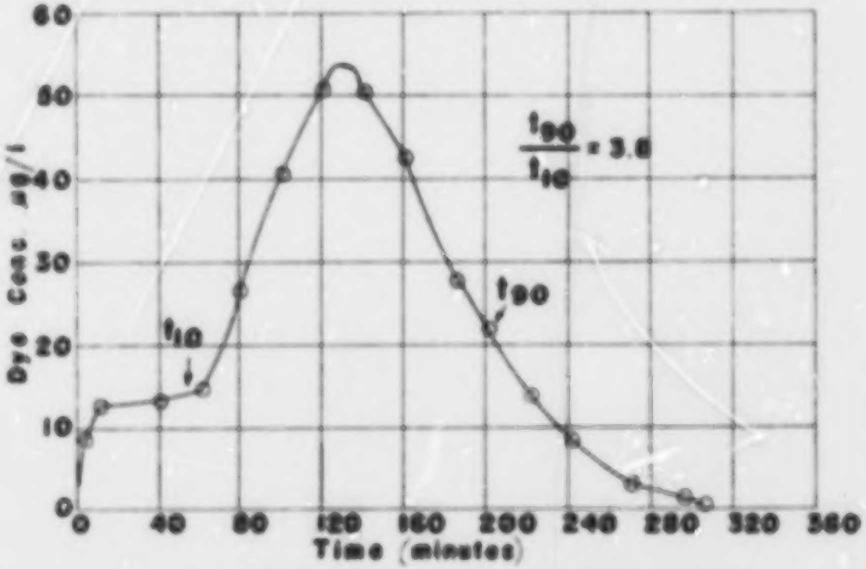
Conclusions and Recommendations

This study indicates that it is not difficult to achieve complete-mixing conditions in an aeration tank. The main problem is designing the tank to achieve maximum utilization of the tank volume for treatment. Recommendations concerning tank design for complete-mixing include the following items:

1. Inlet points - number & type
2. Effluent weir design
3. Diffuser density
4. Air flow
5. Improvements in diffuser placement

Table 1 shows similar n values for both multiple and single inlet arrangements. However, two of the single inlet tests gave washout curves with changing k values. Also runs 14, 15, and 16 exhibited a short duration peak of dye at the start of each test. Although this amounted to only about 2% of the total dye recovered, it does represent a degree of short circuiting. With this in mind, a minimum of 2 inlet points each for primary effluent and return sludge for tanks of a size similar to Freeport is recommended. The inlet points should be located symmetrically on one side of the tank.

The transverse flow pattern to a long sharp crested effluent weir performed very well. Diffusers should not be located directly underneath the effluent channel. The air-water mixture tends to cause periodic uneven flow over the



TRACER CURVE – RUN 12
CONFIGURATION III – PLUG FLOW
Figure 23

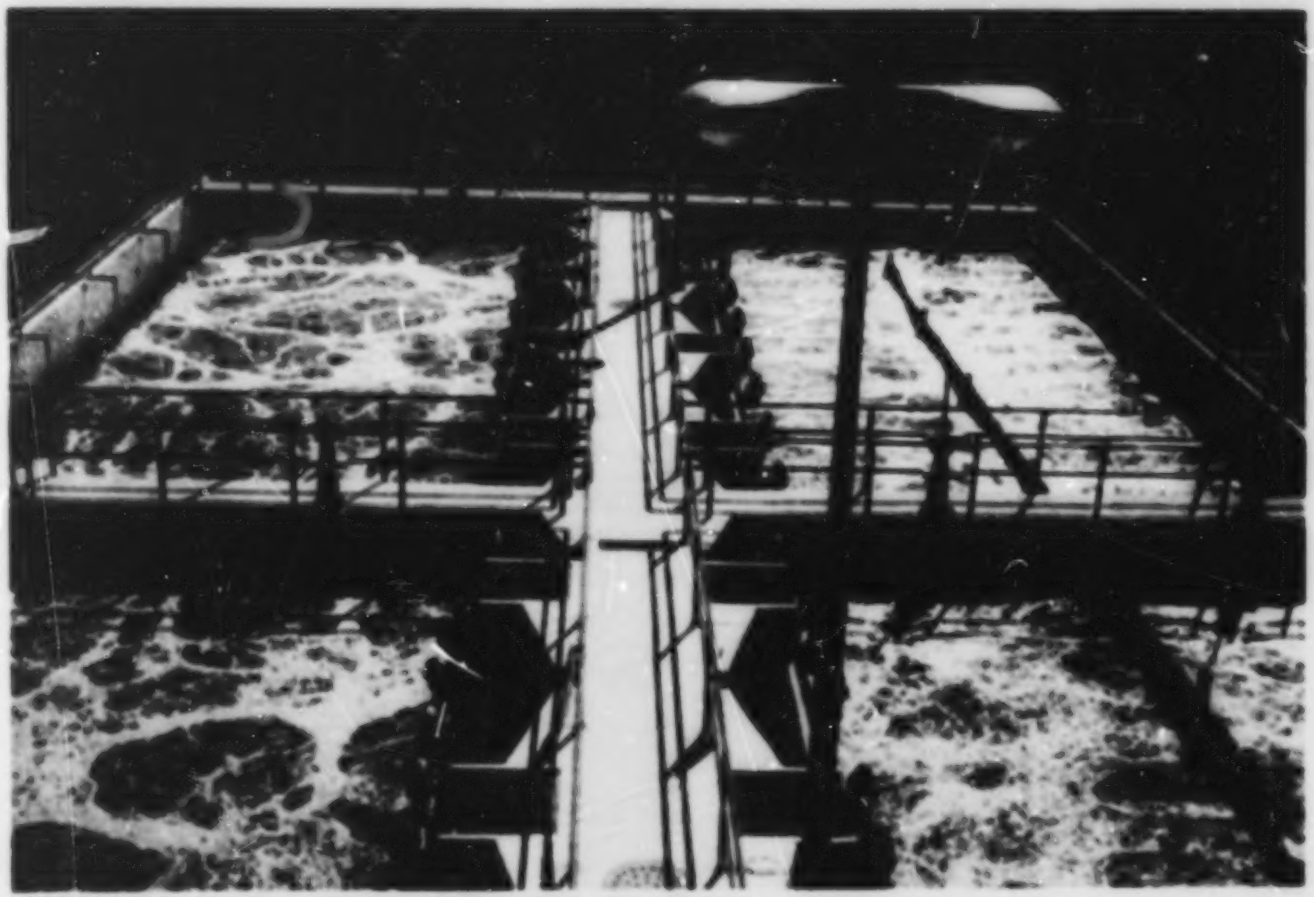
weir.

One final diffuser arrangement was tested in October 1971. The first 7 pairs of diffusers were removed from the inlet side of each header. This resulted in total of 165 diffusers (see the bottom diagram in Figure 5). Flow was as in C4 I shown in Figure 12, i.e., 5 inlet points and transverse flow.

The results of the test are as follows:

$n = 1.12$
$k = .0112$
$\bar{T} = 1/k = 1.5$ hrs
(This compares with a nominal $T = 1.2$ hrs)
% dye recovery = 96%

This data indicates the most effective use of the tank volume in a CM situation of all the arrangements tested.



ACTIVATED SLUDGE SYSTEM OPERATING IN CM MODE I ON THE LEFT AND PF MODE III ON THE RIGHT

FIGURE 24

SECTION V

APPROACH TO EVALUATING BIOLOGICAL PERFORMANCE

Introduction

To evaluate the performance data obtained from the study, several approaches were used. The three basic parameters considered were treatment efficiency, oxygen uptake, and solids production. Equations relating these parameters were taken from analyses developed by W.W. Eckenfelder, Ross McKinney, and H.S. Smith.

Solids production equations were written using Eckenfelder's equations and the active mass balance as described by Smith.

Oxygen uptake constants were developed from oxygen uptake tests and expressed in an oxygen balance as outlined by Smith and an uptake equation presented by McKinney.

Removal efficiencies were related to McKinney's reaction rate constant.

The above analyses are outlined briefly in the following sections.

System BOD Balance

Figure 25 shows a schematic of a CM activated sludge system outlining the balance of ultimate BOD as presented by Smith[9]. The terms used are defined as follows:

Q - influent flow rate
rQ - return sludge flow rate
F_o - PE BOD_L unfiltered
F - FE BOD_L filtered
V_a - aeration tank volume
K_4 - removal rate constant (1/time)
t - nominal detention time in aeration tank

By writing a mass BOD balance around the aeration tank the following equation results.

$$QF_o + rQF = K_4FV_a + QF + rQF$$

$$F_o = \frac{K_4FVa}{Q} + F = F(K_4t + 1)$$

or $$\frac{F}{F_o} = \frac{1}{(K_4t + 1)}$$

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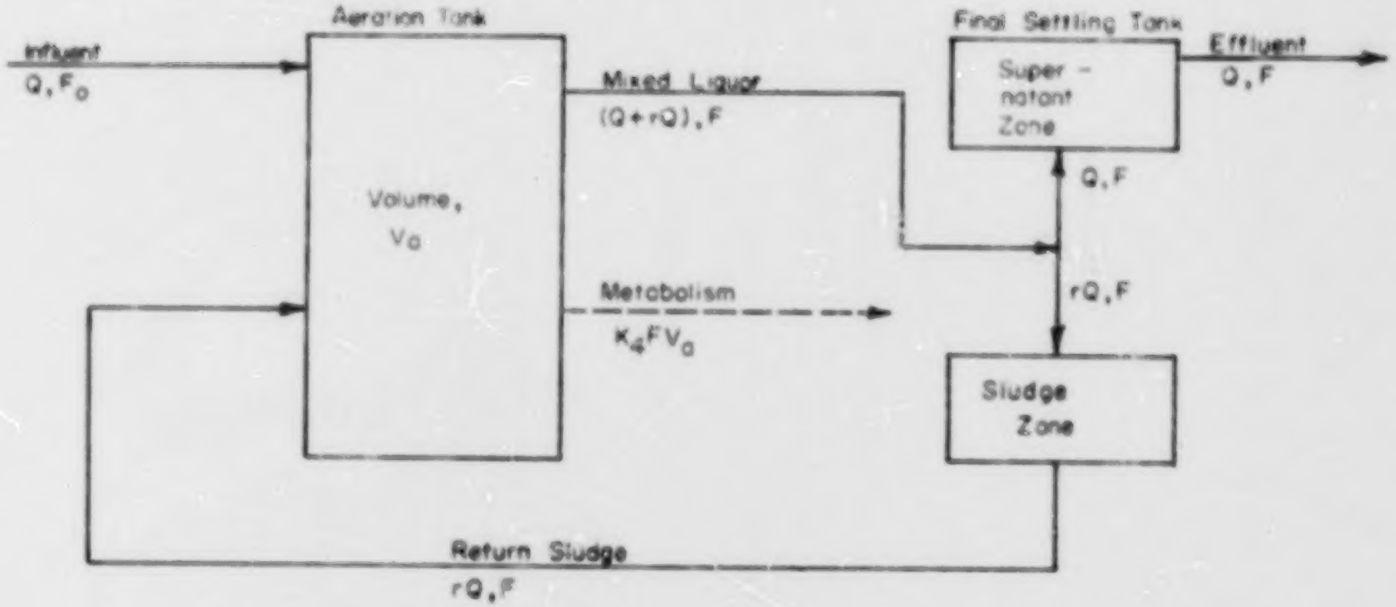

Influent
Q, F_o
Aeration Tank
Volume,
V_a
Mixed Liquor
$(Q+rQ), F$
Metabolism
K_4FV_a
Final Settling Tank
Super - natant Zone
Effluent
Q, F
Q, F
rQ, F
Sludge Zone
Return Sludge
rQ, F

BOD BALANCE

Figure 25

The above analysis makes the assumption that K_4 is a constant and not time dependent.

System Oxygen Balance

Figure 26 shows a complete oxygen balance for a CM system as described by Smith[9]. The terms used in this figure are defined as:

Q = influent flow rate
rQ = return sludge flow rate
F_o = FE BOD_L unfiltered
F = FE BOD_L filtered
V_a = aeration tank volume
M_a = active mass in mixed liquor (oxygen units)
M_{ar} = active mass in return sludge (oxygen units)
K_e = endogenous rate constant (1/time)
K_s = synthesis rate constant (1/time)
ΔM_a = net change in M_a of flow $(Q + rQ)$

Then, by writing an oxygen mass balance around the aeration tank:

$$(1) \quad QF_o + rQ\ (F + M_{ar}) = (Q + rQ)\ (F + M_a) + K_eM_aV_a + K_sFV_a$$

The active mass balance around the aeration tank can also be expressed by:

$$(2) \quad \overset{\text{Mass In}}{(Q + rQ)\ \Delta M_a + rQM_{ar}} = \overset{\text{Mass Out}}{(Q + rQ)\ M_a}$$

or

Equation(1)can be rewritten:

$$QF_o + rQF + rQM_{ar} - (Q+rQ)F - (Q+rQ)M_a = V_a(K_sM_a + K_eF)$$

substituting equation(2)

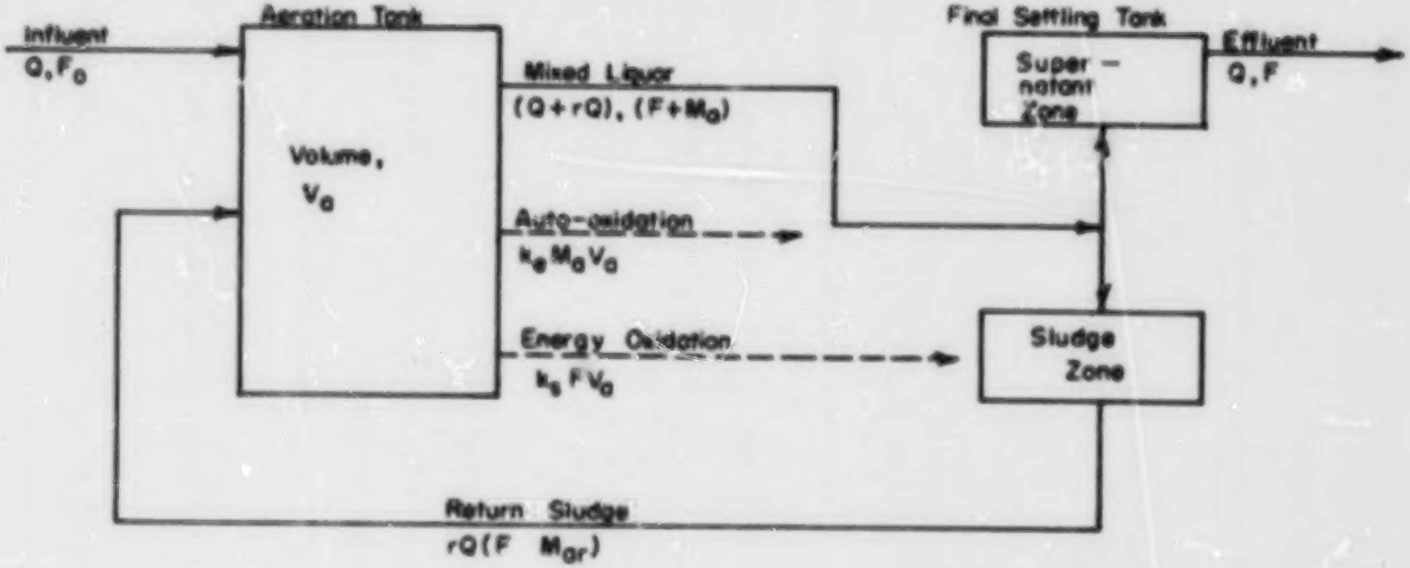
$$QF_o + rQF + rQM_{ar} - (Q+rQ)F - (Q+rQ)\ \Delta M_a - rQM_{ar} = V_a(K_sM_a + K_eF)$$

$$QF_o + rQF - QF - rQF - (Q+rQ)\ \Delta M_a = V_a\ (K_sM_a + K_eF)$$

dividing by Q

$$(F_o - F) - (1+r)\ \Delta M_a = t\ (K_sM_a + K_eF) \quad (3)$$

Equation (3) expresses concentration of food supplied in terms of oxygen units (left hand term) to concentration of oxygen used for synthesis and endogenous respiration (right hand term).


Influent
Q, F_o
Aeration Tank
Volume,
V_a
Mixed Liquor
$(Q+rQ), (F+M_a)$
Auto-oxidation
$k_e M_a V_a$
Energy Oxidation
$k_s F V_a$
Final Settling Tank
Super-natant Zone
Sludge Zone
Effluent
Q, F
Return Sludge
$rQ(F \quad M_{ar})$

OXYGEN BALANCE

Figure 26

The rate of oxygen consumed can then be related to equation (3) by the equation as expressed by McKinney[10]:

$$(4) \quad \frac{do}{dt} = K_s M_a + K_e F$$

System Active Mass Balance

To complete the analysis of the CM system a complete flow diagram for active mass is shown in Figure 27. Definition of terms is as follows:

Q - influent flow rate
rQ - return sludge flow rate
wQ - waste sludge flow rate
F - FE BOD_L filtered
Va - aeration tank volume
M_a - active mass in mixed liquor
M_{ar} - active mass in return sludge
M_{ae} - active mass in final effluent
K_e - endogenous rate constant
K_s - synthesis rate constant

An active mass balance around the aeration tank yields the following equation:

$$(5) \quad \cancelto{0}{QM_{ai}} + rQ\, M_{ar} + K_s F V_a = (Q+rQ)\, M_a + K_e M_a V_a$$

Tests show M_{ai} to be zero, so this term can be eliminated. Equation (5) can be rearranged:

$$rQM_{ar} - (Q+rQ)M_a = K_e M_a V_a - K_s F V_a$$

or

$$(Q+rQ)M_a - rQM_{ar} = K_s F V_a - K_e M_a V_a$$

Now, a balance around the final settling tank results in the equation:

$$(6) \quad (Q+rQ)M_a = QM_{ae} + rQM_{ar} + wQM_{ar}$$

Substituting (6) into (5) yields:

$$QM_{ae} + rQM_{ar} + wQM_{ar} - rQM_{ar} = V_a\, (K_s F - K_e M_a)$$

dividing through by Q:

$$(7) \quad M_{ae} + wM_{ar} = t(K_s F - K_e M_a)$$

Eckenfelder Solids Balance

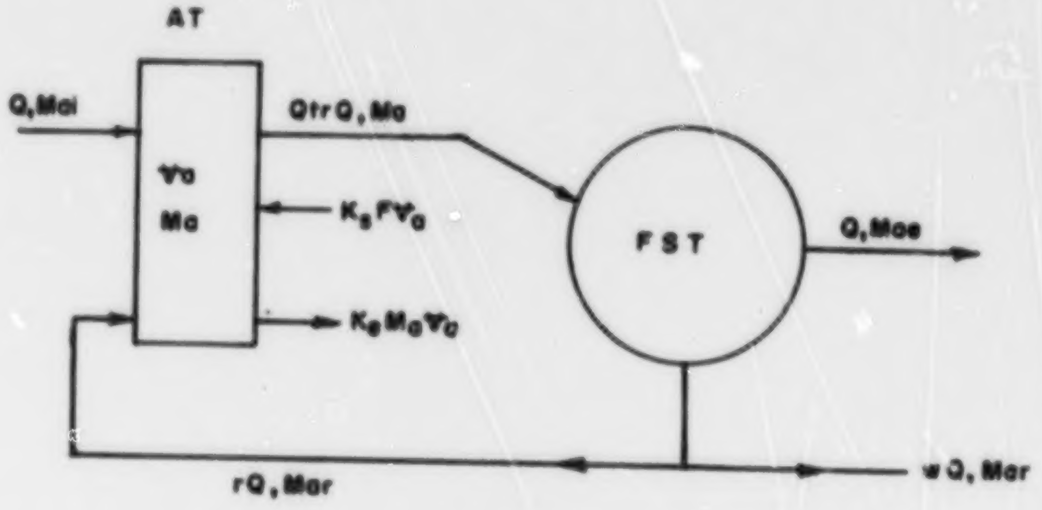
The following terms are presented by Eckenfelder[11] to develop a solids production equation.

ΔX - sludge production per day
V - volume of aeration basin
Q - flow
a - fraction of substrate converted to new cells
S_r - substrate removed
b - fraction per day of VSS oxidized
X_a - MLVSS
X_o - influent SS
X_e - effluent SS

The above terms are related by the following equation:

(8) $\Delta X = QX_o + aSrQ - (bXaV + QX_e)$

To determine the constants a and b a plot is made of $\Delta X/X_a$ versus S_r/X_at. The slope of the curve is a and the y - intercept is b.


AT
Q,Mai
Q+rQ, Ma
Va
Ma
K$_s$ F V$_a$
K$_e$ M$_a$ V$_a$
FST
Q,Mae
rQ, Mar
wQ, Mar

ACTIVE MASS BALANCE

Figure 27

SECTION VI

CONSTANT DETENTION TIME STUDY

To determine the performance of the CM system under a range of detention times, eight different constant influent flow rates were applied to aeration tanks 1 & 3. The flow pattern used for all the periods was transverse flow with one center inlet point. An adjustable stop gate was installed in each center inlet point, and the operators adjusted these gates to maintain a constant flow as measured by the final settling tank Parshall flumes. The remainder of the plant flow went through aeration tanks 2 & 4, so these tanks essentially acted as a buffer system.

The following testing procedures were carried out for this phase of the study and also continued through the parallel operation phase.

1. Composite samples of primary and final effluent were taken on a daily basis using the samplers described in the introduction section and located as shown in Figure 1. Samples were collected from the automatic units and refrigerated every two hours. Lab analyses were run as indicated below:

   COD Filtered & Unfiltered

   BOD_5 Filtered & Unfiltered

   TSS & VSS

   NH_3

   NO_3

   (Details of laboratory analyses are included in the appendix)

2. Samples of aeration tank mixed liquor and return sludge were taken every 8 hours and the following determinations were made:

   MLSS & MLVSS (solids procedure described in appendix)
   RS SS & RSVSS

   Aeration Tank D.O. & Temperature

   Sludge Settled, 30 Minutes

3. O_2 uptake tests were performed on samples of mixed liquor, return sludge, and final effluent periodically during the testing period (These are discussed in detail in the parallel operation section of the report).

Table 4 is a summary of the operating data for the CM system for the 8 constant T periods. An attempt was made to hold the loading in terms of lb BOD_5/lb MLVSS constant throughout the tests. With this in mind, MLSS values were increased as the constant influent flow rate increased. However, this procedure was not entirely successful as evidenced by the increasing F/M ratios with increasing flow rates. MLVSS values ranged from 1850-3340 mg/l with an average of 2760 mg/l.

The parameters used to evaluate the system performance included the following:

1. % removal BOD_5 PE UF to FE F & PEUF - FE UF

2. % removal COD PE UF to FE F & PEUF - FEUF

3. FE BOD_5 values

BOD_5 Removal

The upper graphs in Figures 28 & 29 show the removal percentages from primary effluent to final effluent for the total and soluble 5-day BOD. The most significant point about this data is that the removal percentages did not decrease appreciably from the 90% range for soluble and the 80-90% range for the total until the 1 hour nominal detention time period.

The soluble data indicates that organic removal in the aeration tanks is almost constant until the mean residence time drops below 1 hour. Actually, the mean residence time was even less than 1 hour according to the results of the dye tracer tests.

Total BOD_5 began to drop off at around 2.5 hours. This can be attributed to reduced performance of the final settling tanks at this flow rate (Refer to section on FST performance). At 2.5 hours nominal T, hydraulic loading on FSTs would be 500 gpd/ft^2.

COD Removal

The COD data shown in Figures 28 & 29 corroborate the conclusions of the preceding section. COD removals are

SUMMARY OF CONSTANT DETENTION TIME STUDIES

Table 4

Time Period	Qe mgd	T (hrs)	Primary Effluent						Final Effluent					
			BOD		COD		SS	VSS	BOD		COD		SS	VSS
			F	UF	F	UF			F	UF	F	UF		
1/1/71 1/31/71	1.30	4.9	82	135	99	287	150	126	14	19	44	70	27	22
2/1/71	1.24	5.1	77	127	98	245	115	92	10	16	40	59	26	25
2/13/71 2/17/71	1.54	4.1	95	135	103	265	127	104	13	13	41	51	16	15
3/12/71 3/17/71	1.59	4.0	39	90	66	208	199	124	7	17	44	54	18	14
3/26/71 4/5/71	2.08	3.0	33	78	67	187	107	80	8	13	39	57	29	23
4/16/71 4/22/71	2.84	2.2	81	122	72	206	126	94	10	15	43	60	36	28
4/24/71 5/9/71	2.85*	1.1	104	156	72	226	161	120	14	33	44	58	27	22
5/10/71 5/31/71	3.53†	0.9	58	121	72	204	104	79	15	35	46	73	29	25

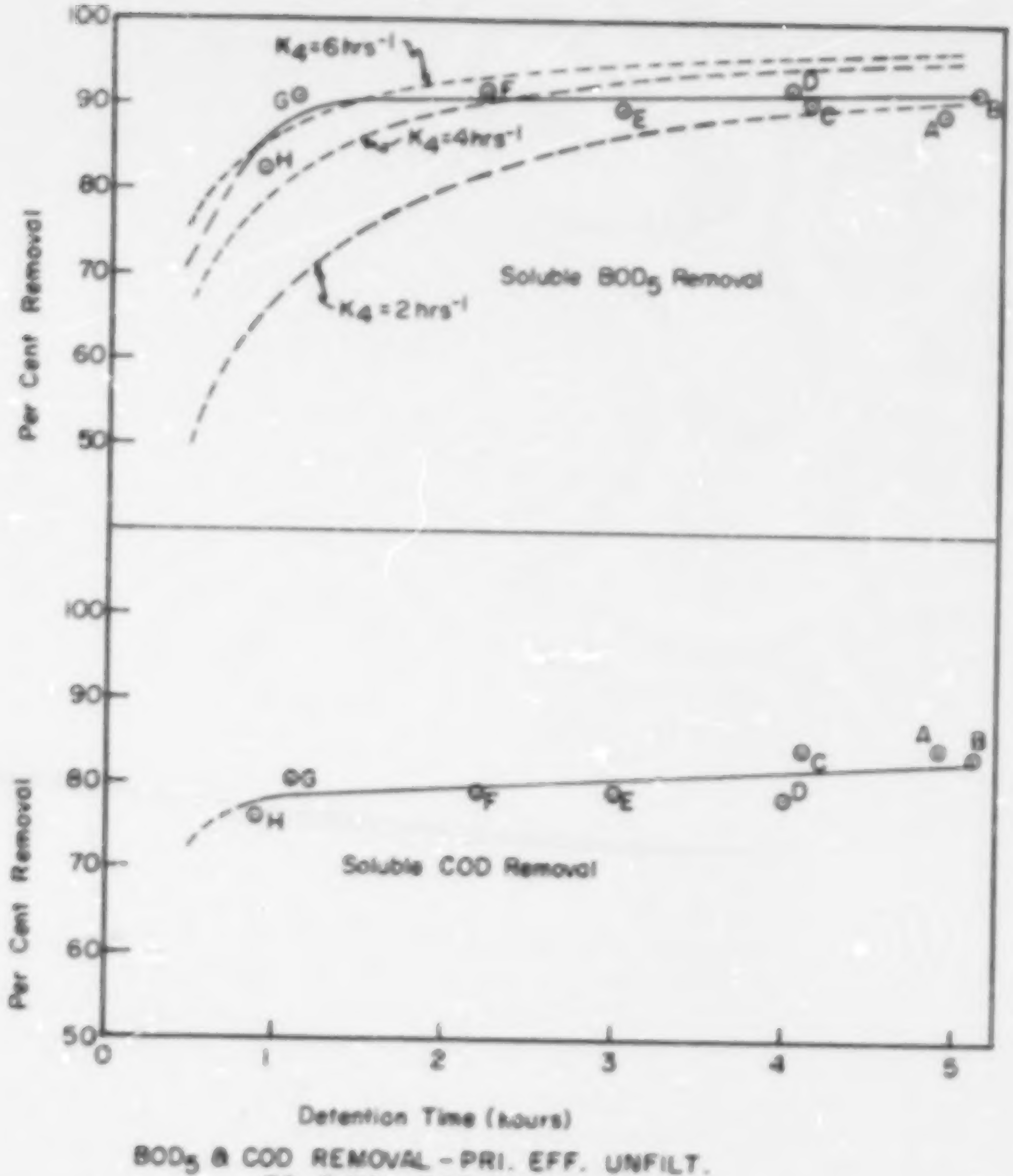
* To only 1 A.T.

† Avg. for diurnal flow variation

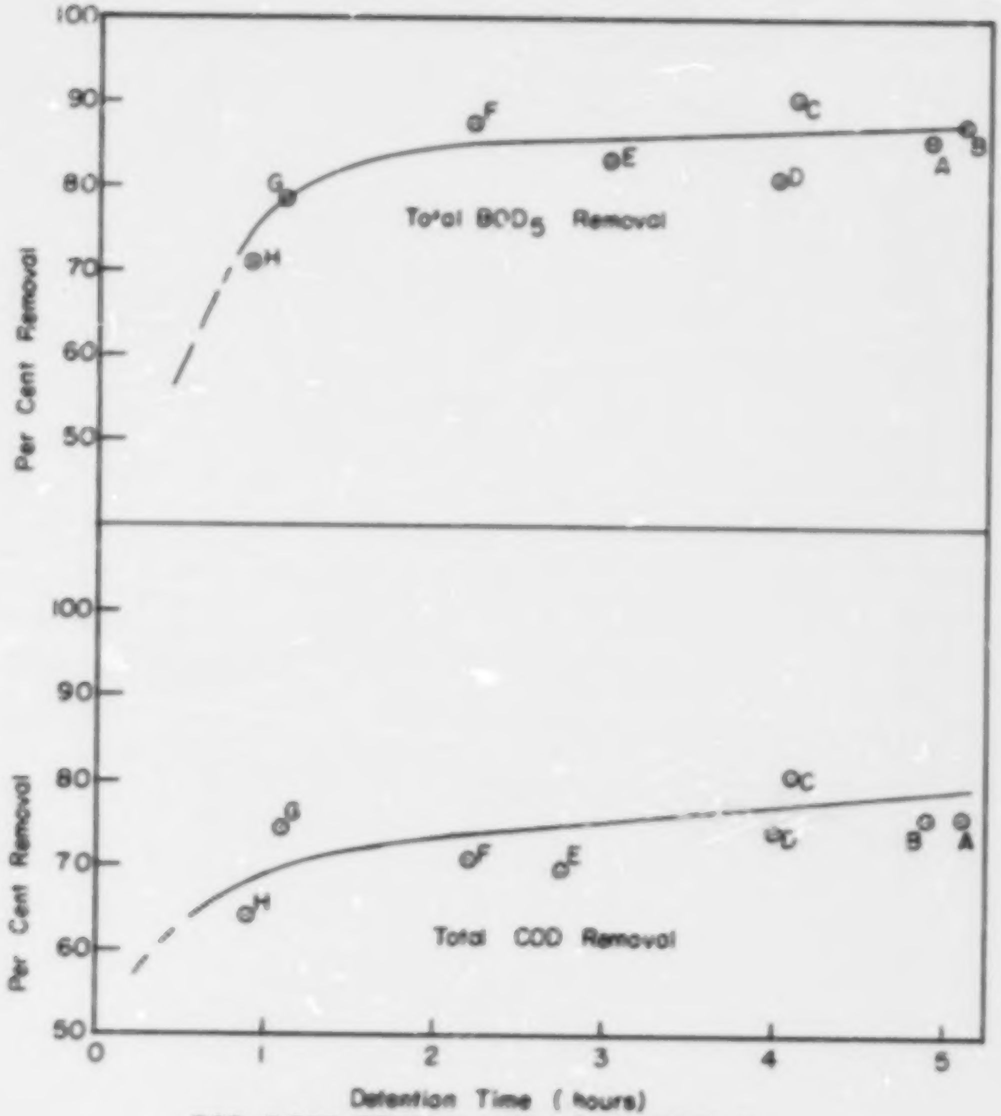
SUMMARY OF CONSTANT DETENTION TIME STUDIES

able 4 cont'd

	Per Cent	Removal					
Time Period	PE-BOD (UF) to FE BOD (F)	PE-COD (UF) to FE COD(F)	SS	VSS	F/M $\frac{\text{lb BOD}_5}{\text{lb MLVSS}}$	Mixed Liquor Temp. (°C)	
1/7/71 1/31/71	89.6	84.7	82.0	82.5	0.26	14	A
2/1/71 2/12/71	92.1	83.7	77.4	72.8	0.19	13	B
2/13/71 2/17/71	90.4	84.5	87.4	85.6	0.24	13	C
3/12/71 3/17/71	92.2	78.8	91.0	88.7	.21	11	D
3/26/71 4/5/71	89.7	79.1	73.0	71.2	.33	12	E
4/6/71 4/22/71	91.8	79.1	71.4	70.2	.51	14	F
4/24/71 5/9/71	91.0	80.5	83.2	81.7	1.00	15	G
5/10/71 5/31/71	87.6	75.0	72.1	68.4	1.14	16	H



BOD$_5$ & COD REMOVAL - PRI. EFF. UNFILT.
TO FINAL EFF. FILT.
Figure 28



BOD$_5$ & COD REMOVAL - PRI. EFF. UNFILT TO FINAL EFF. UNFILT.

Figure 29

in general lower than BOD_5 removals, but again there is a gradual decrease in efficiency until 1 hour is reached for soluble values and a more pronounced decrease in efficiency with decreasing T for total COD values. It should be noted, however, that the soluble COD efficiency decrease around 1 hour is not as marked as the corresponding one for BOD_5.

Final Effluent BOD_5 Values

Though per cent removal is a good indicator of system performance for comparison purposes, most regulatory agencies are basing evaluation of plant operation on effluent criteria. The data plotted in Figure 30 gives a good indication of how the CM system would perform if an effluent BOD_5 value of 20 mg/l is chosen as a lower limit. It shows the percent of time that the system would be in compliance.

On this basis, the break point is between 2 and 3 hours. This is probably the best criteria to use for purposes of aeration tank design.

Removal Rate Constant

A common method[10] of characterizing the removal rate is the formula:

$$\frac{F}{F_o} = \frac{1}{K_4 t + 1}$$

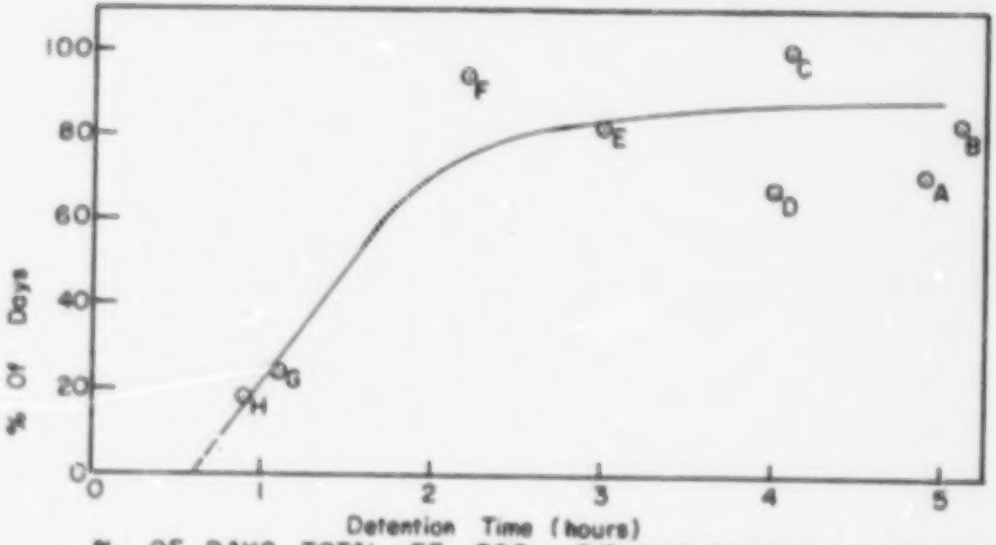
, where F = Effluent soluble BOD
F_o = Influent soluble BOD
t = Detention time
K_4 = Rate constant

Figure 28 shows 3 removal curves corresponding to $K_4 = 2$, 4, and 6 hrs^{-1} superimposed on the actual measured data. The actual curve is very flat past 2 hours detention time, indicating that the apparent K_4 value is not a constant but changes with T.

As an approximation $K_4 = 4\ \text{hrs}^{-1}$ is the best value for following the measured data. It can be said that the removal curve is bounded by $K_4 = 2\ \text{hrs}^{-1}$ and $K_4 = 6\ \text{hrs}^{-1}$.

Other Analysis Of Data

Several other analyses were made on the constant detention data, but it is directly connected with the parallel operation phase so it is all presented in that section. The analyses include interpretations of the O_2 uptake data and development of prediction equations for air requirements and solids production.



% OF DAYS TOTAL FE BOD_5 DID NOT EXCEED 20 mg/l

FIGURE 30

SECTION VII

PARALLEL OPERATION

Introduction

From the results of the tracer studies, the flow patterns shown in Figure 1 were adopted for the full scale PF versus CM parallel operation. Specifically, PF consisted of 2 end influent points and a baffled aeration tank system 230' long by 12.5' wide (L/W = 18.4:1) with a 10 foot end effluent weir. The CM system consisted of two transverse flow tanks with 5 inlet points and side effluent weirs. Diffuser placement was as shown in the lower diagram in Figure 6 for PF and as shown in the 198 diffuser diagram in Figure 5 for CM. Return sludge entered at the influent end of the PF system (See Figure 4). Four side inlet points were used for return sludge on the CM side. During certain periods of the parallel study, there occurred some variations in the above configurations. Significant changes will be described where they apply in the discussion.

Parallel operation was conducted from June to November 1971. For the majority of time the flow was split equally between the two sides. The basic plan was to compare performance between the sides with equal loads under different concentrations of MLSS. Sample collection was continued as in the constant T phase of the project.

Discussion of parallel operation is divided into four parts:

a) Operating data and comparison of parallel performance
b) Oxygen uptake studies
c) Solids production studies
d) Final settling tank performance

Operating Data and Comparison of Parallel Performance

The parallel operation phase was divided into nine periods designated I-Q in Table 5. Periods were chosen to group numbers of days together during which MLSS values were similar and flow patterns were identical for each side of the plant. Thus, for each period the loading (F/M) was about the same for CM and PF and the nominal detention time was the same for each (Refer to Table 5).

To compare performance the following parameters were chosen:

a) Per cent removal of 5 day soluble BOD from primary effluent to final effluent, i.e., removal based on the difference between unfiltered BOD_5 in the primary

REMOVAL EFFICIENCIES
PARALLEL OPERATIONS

Table 5

Date	Period	BOD Removal PEFF-FEFF (%)		COD Removal PEFF-FEFF (%)		T (hrs)	F/M lb. BOD$_5$ / lb. MLVSS day	
		CM	PF	CM	PF		CM	PF
6/9-6/16 & 6/20-6/22	I	84.2	81.0	78.4	79.0	3.1	0.43	0.38
7/1-7/11 & 7/26-7/31	J	89.4	90.1	75.5	78.7	3.3	0.82	0.75
7/12-7/25	K	92.0	90.4	82.0	82.0	3.2	0.57	0.57
8/1-8/9	L	92.1	92.1	79.2	79.2	3.1	0.55	0.54
8/10-8/22	M	88.9	90.0	82.3	80.5	3.1	0.44	0.53
8/23-9/6	N	90.5	89.3	75.0	74.3	3.2	0.37	0.41
9/7-9/12	O	90.0	88.5	77.5	78.7	3.1	0.27	0.28
9/20-9/30	P	90.9	91.9	75.9	78.7	3.2	0.31	0.33
10/1-10/10	Q	89.6	88.4	85.7	80.2	3.5	0.27	0.26

FINAL EFFLUENT BOD$_5$ EVALUATION
PARALLEL OPERATION

Table 6

Period	Avg. FE BOD$_5$		Days 20 mg/l Exceeded		Total Days	% time not exc.		Avg. MLVSS	
	CM	PF	CM	PF		CM	PF	CM	PF
I	27	22	5	5	11	54.5	54.5	1714	1933
J	34	34	9	11	17	47.1	35.3	1100	1198
K	28	19	5	3	14	64.3	78.6	1272	1235
L	16	12	1	0	9	88.9	100	1023	1045
M	15	14	3	0	12	75.0	100	1375	1151
N	24	24	6	6	15	60.0	60.0	1360	1510
O	15	22	0	2	6	100	66.7	1990	1515
P	17	15	1	1	11	90.9	90.9	2330	2010
Q	16	14	3	3	10	70.0	70.0	2340	2240

effluent and filtered BOD_5 in the final effluent. This essentially evaluates the performance of the aeration tank system, but does not indicate solids removal effectiveness in the final settling tank.

b) Similarly, percent removal of soluble COD from PE to FE

c) Total BOD_5 and COD concentration in each final effluent for a given applied concentration in the primary effluent

d) Suspended solids concentrations in each FE

e) Finally, as a measure of the ability of each system to meet a set effluent criteria, an evaluation was made of the percent of time during which each system exhibited a final effluent concentration below 20 mg/l BOD_5.

Figure 31 shows the comparisons of removal efficiencies of soluble BOD_5 and COD. The average for each period is plotted as a single point at the midpoint of the period. This data does not indicate a significant difference in removal efficiency between the two systems. Generally, efficiencies remained around 90% for BOD_5 and 80% for COD.

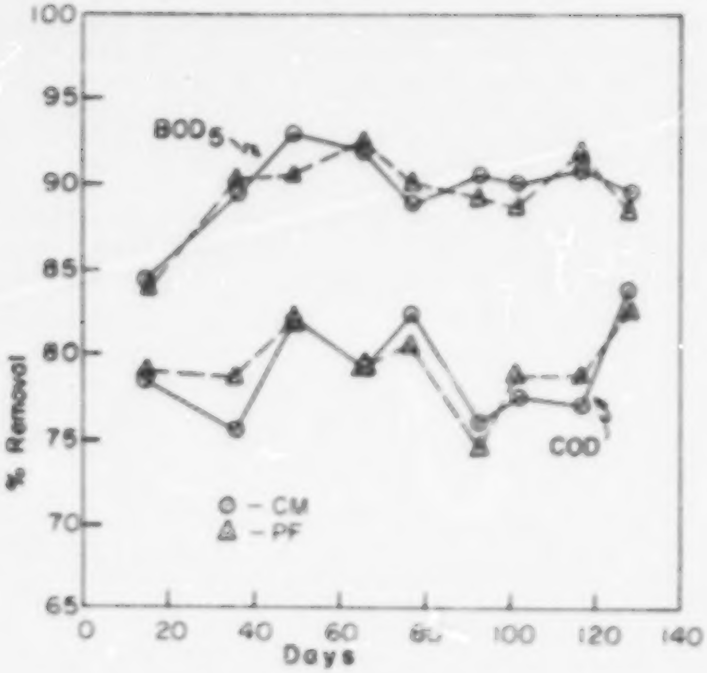
The ability of each system to meet a 20 mg/l effluent standard is outlined in Table 6. Though PF showed some superiority during the middle periods CM was better during some others.

The actual effluent averages are shown in Table 6. On the basis of an overall weighted average, BOD_5 for PF was 21 mg/l and for CM was 23 mg/l. The most significant point of this evaluation is that neither system met the 20 mg/l criteria 100% of the time.

Average values of BOD_5 and COD for PE and FE are shown graphically in Figure 32. All FE BOD_5 values are very similar for CM and PF. During period K the FE COD for PF did come out considerably higher than CM. Other than this COD values were also all very similar.

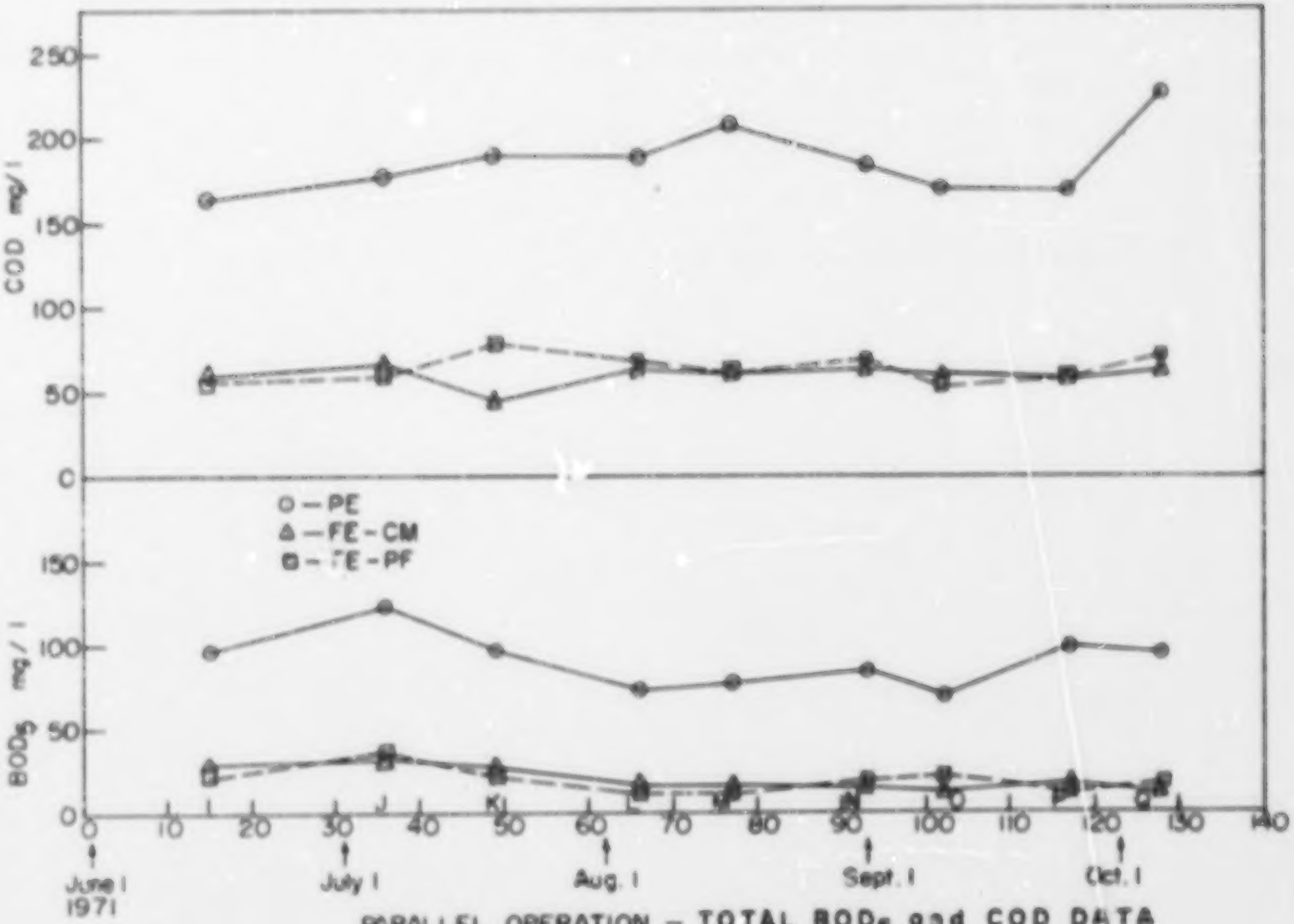
The solids data is outlined in Figure 33. There was a trend towards higher effluent solids in the PF for periods K and L.

It is difficult to make any definitive statement concerning removal efficiency as it relates to organic loading. The F/M ratios are recorded in Table 5. There is a trend towards slightly higher removal efficiency with a decrease in F/M (See periods I, J, O, and Q). One can not clearly distinquish between the 2 modes of operation.



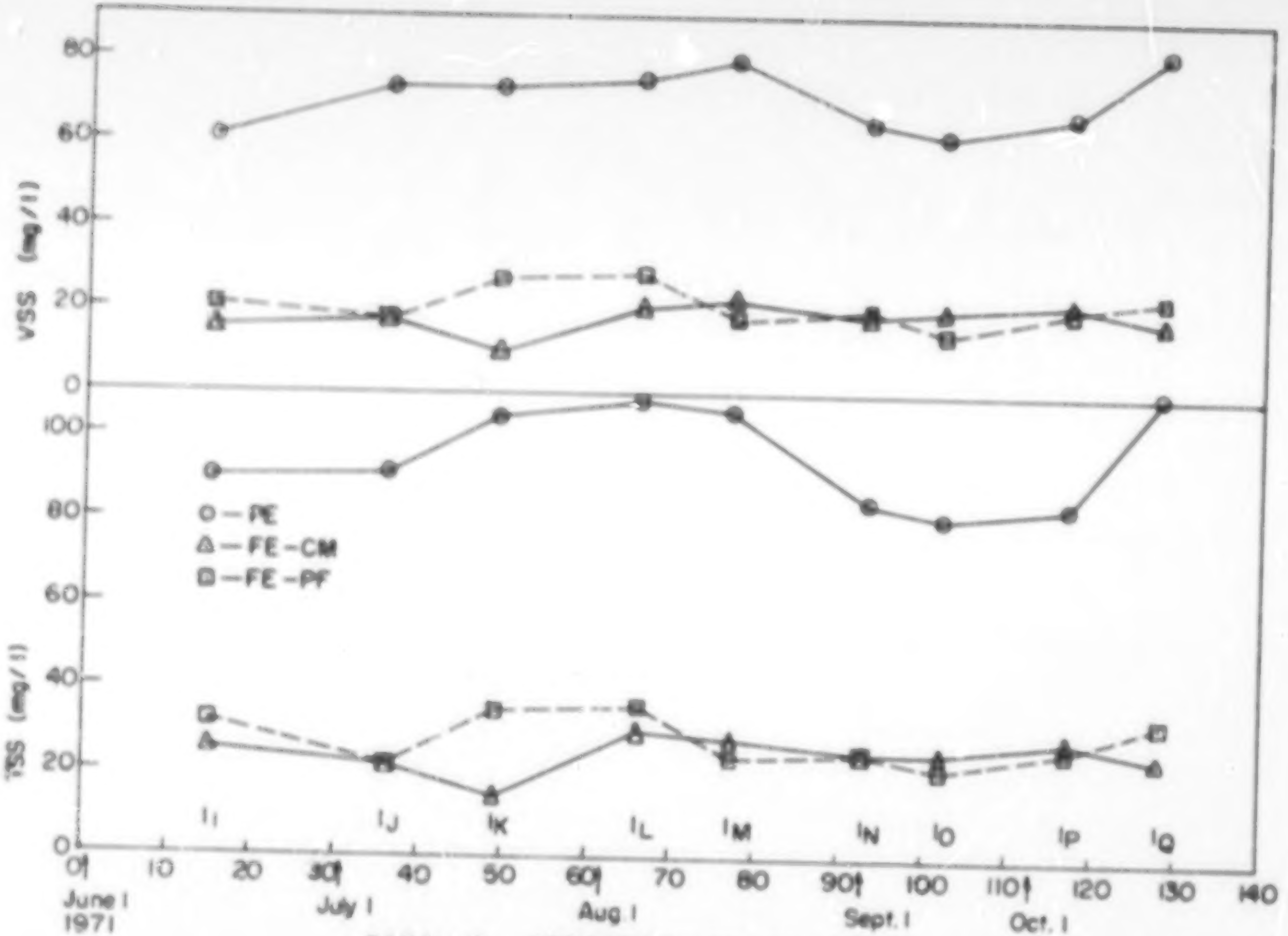
PARALLEL OPERATION
REMOVAL EFFICIENCIES (SOLUBLE)

Figure 31



PARALLEL OPERATION – TOTAL BOD$_5$ and COD DATA

Figure 32



PARALLEL OPERATION - SOLIDS DATA
Figure 33

Similarly, Figure 34 shows a plot of removal versus lb BOD_5/1000 ft^3. Here again, there is no marked difference between CM and PF.

Finally, Figure 35 is a synopsis of soluble BOD_5 removal efficiencies for varying detention times for both CM and PF. This plot is referred to in the design analysis section of the report.

Possible Effect of Waste Characteristics

It must be recognized that the Freeport sewage is a relatively weak domestic-commercial waste with only limited industrial contributions. The total BOD_5 of the aeration tank influent seldom exceeded 150 mg/l and often was less than 100 mg/l.

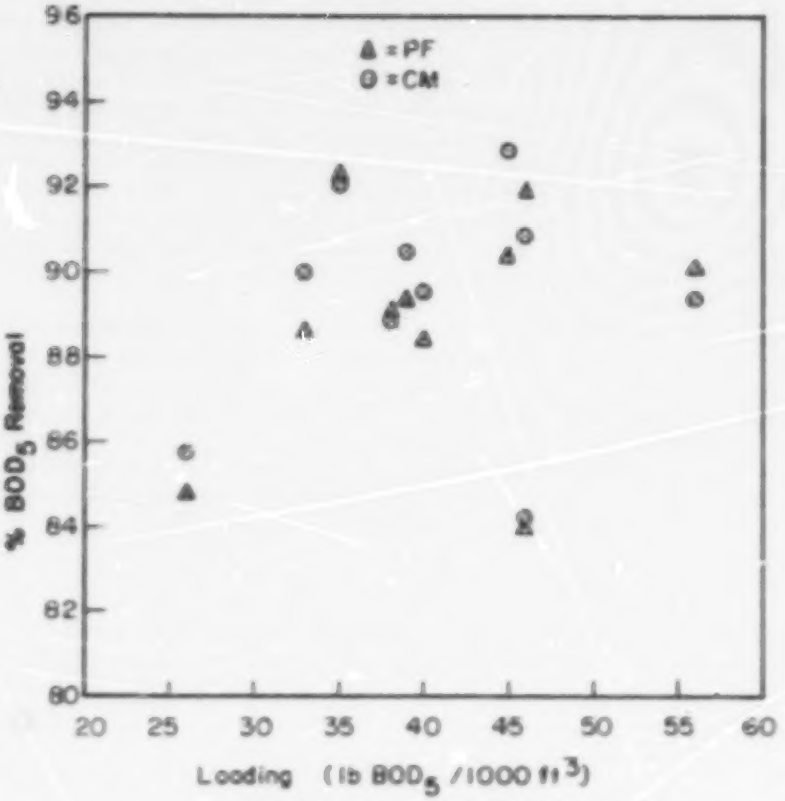
An inherent difference in biological environment in CM and PF systems is the uniform substrate composition throughout the CM system compared with the variable concentration in the PF system. At the head end of the PF system the influent is dispersed in only a small part of the tank volume into which return sludge is added. Not only is the F/M ratio high in the head end of the PF system but the return sludge organisms face another shock situation in as much as they are coming from the region of the system in which substrate concentration is not only the lowest but of different character than that presented by the influent.

The effect of these differences in the CM and PF systems could have been masked or minimized in this study due to the weak waste. This tends to minimize the environmental differences posed by the two systems. This would be particularly true when the systems are operated well within their capabilities. It is expected that the effect of differences might be more pronounced when the systems are operating near the limit of their capabilities.

Evaluation of Air Requirements

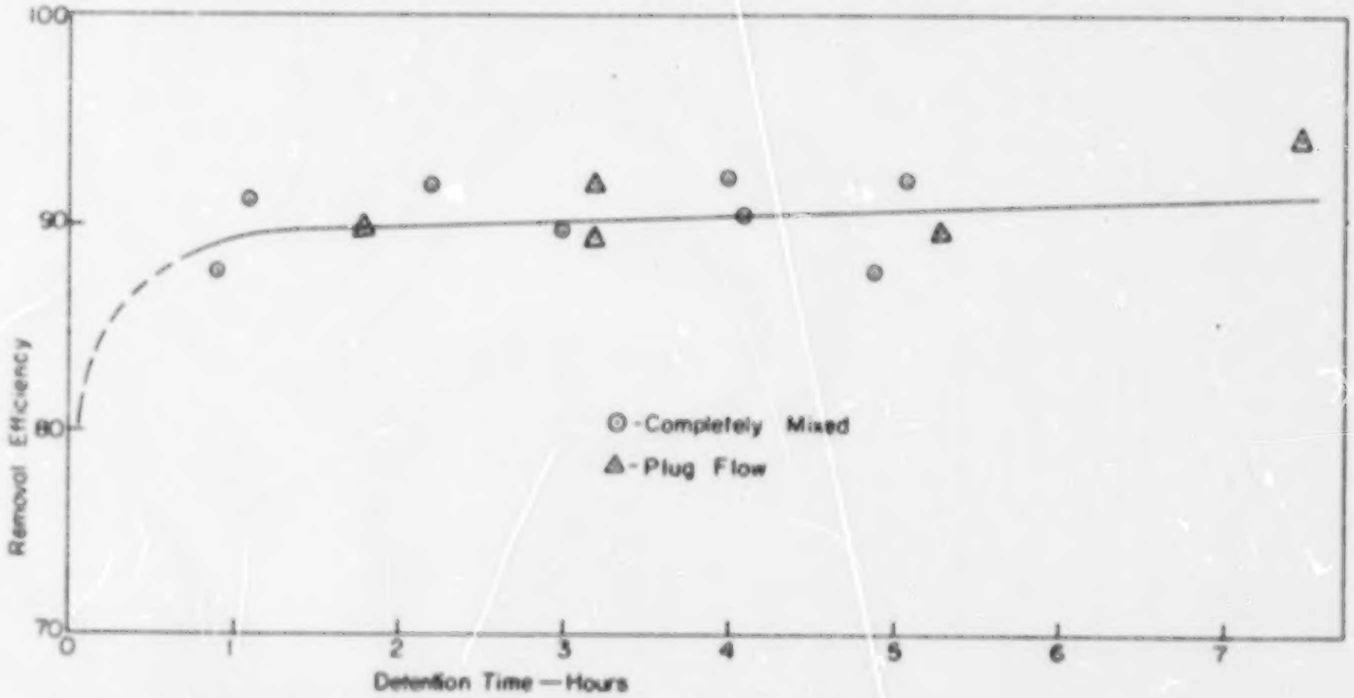
The following evaluations were made to describe the air requirements of each system. Each evaluation will be discussed in detail separately. A discussion of how the overall results apply to design is then presented:

a) O_2 uptake tests on completely-mixed system
b) O_2 uptake tests on plug-flow system
c) 24 hour time-dependent air flow versus DO tests
   on both systems
d) O_2 transfer efficiency tests on both systems
e) Parallel O_2 uptake tests on a side by side basis



PARALLEL OPERATION
VOLUMETRIC BOD LOADING

Figure 34



CONSTANT T, SOLUBLE BOD_5 REMOVAL EFFICIENCIES
Figure 35

O_2 Uptake Tests on CM System

From December 1970 to June 1971, a series of O_2 uptake tests were run on samples of mixed liquor, return sludge and final effluent from the CM side of the plant.

Each test consisted of removing a given sample from the flow, placing it in a sealed container under controlled temperature conditions and measuring the change in DO versus time with the use of a probe placed in the container (A detailed description of the lab and analytical procedure are included in the Appendix of the report).

Each test was designed to determine the following parameters:

1. Initial O_2 uptake rate
2. Energy respiration constant K_s (hrs^{-1})
3. Endogenous constant K_e (hrs^{-1})
4. Active mass Ma(mg/l)

Table 7 is a summary of the results of these tests(All results of these tests are reported at a temperature of 20°C). To utilize this data in an active mass (Ma) balance equation, averages of the pertinent parameters (K_s, K_e, and Ma/VSS were determined and are presented in Table 8.

The constants K_s and K_e for use in predicting O_2 uptake and solids production were determined in the following manner:

1. On the basis of the average data in Table 8 the active mass to volatile suspended solids ratios were set at Ma/VSS = Mar/VSS = 0.26
   Mae/VSS = 0.40

   The endogenous respiration constant K_e was set at,

   $K_e = .026\ hrs^{-1}$

2. Average operating data for CM was summarized for 8 constant detention time periods (See Table 9) and 7 normal diurnal-variation periods (See Table 10).
3. For each period, the equation $Mae + wMar = T\ (K_sF - K_eMa)$ was solved for K_s using $K_e = .026$ and the appropriate operating data.
4. The average K_s for the set of periods was then

Summary of O_2 Uptake Data - Completely Mixed

Table 7

Date	Sample Location	Ks (hrs^{-1}) (O_2Units)	Ke* (hrs^{-1}) O_2Units	F mg/1	Ma (mg/1) (O_2Units)	Ma (mg/1) (solids units)	VSS (mg/1)	$\frac{Ma}{VSS}$	Total O_2 Uptake @ t =0 (mg/1.hr)
12 8 70	AT 1 RL	1.03	.0572	22.0	594	418	1310	0.32	44.1
1/11/71	AT 1 RL	-	.0214	-	975	686	2480	0.28	41.0
[illegible]	AT 1 RL	-	.0254	-	833	587	2360	0.24	35.0
[illegible]	AT 1 RL	-	.0092	-	2411	1698	2530	0.67	-
[illegible]	AT 1 RL	1.48	.0221	6.1	1175	827	3100	0.28	55.0
[illegible]	AT 1 RL	0.72	.0131	13.1	1240	874	2570	0.34	29.4
[illegible]	AT 1 RL	0.83	.0152	7.9	745	524	2160	0.24	16.4
[illegible]	AT 1 RL	2.08	.0472	7.2	571	402	1820	0.22	41.0
1/27/71	AT 1 RL	1.18	.0235	23.4	990	697	3230	0.22	56.0
3/2[illegible]/71	AT 1 RL	1.03	.0442	12.1	260	183	980	0.19	24.0
4/15/71	AT 1 RS	-	.0159	-	3024	2130	6840	0.31	-
[illegible]/71	AT 1 RS	-	.0204	-	2995	2109	7130	0.30	-
3/17/71	AT 1 RS	-	.0193	1.22	1790	1260	6500	0.19	-
3/25/71	PST IPE	-	.0412	-	6.4	4.5	11	0.41	
5/5/71	AT 1 RS	-	.0228	1.03	4605	3243	11720	0.25	-

---

*K_e when used in equations will be expressed in solids units, i.e.,(K_e above) x 1.42 **(1.42 is the assumed ratio, O_2 units: solids units)**

Summary of O_2 Uptake Constants

Completely-Mixed

Table 8

Type of Sample	No. of Tests	Avg. K_s (hrs^{-1}) O_2 Units	Avg. K_e ($hrs.^{-1}$) O_2 Units	Avg. Ma/VSS
Mixed Liquor	10	1.19*	0.026	0.26 †
Return Sludge	4	1.12††	0.020	0.26
Final Effluent	1	-	0.041	0.41

* Average of 7 Tests

† Omitting 1 High Value of 0.67

†† Average of 2 Tests

ACTIVE MASS BALANCE
OPERATING DATA

January - May 1971

Table 9

Period	FE-VSS (mg/l)	Mac (mg/l)	ML VSS (mg/l)	Ma (mg/l)	RS VSS (mg/l)	Mar (mg/l)	w	t (hrs)	FE-BOD$_5$ (mg/l)	F (mg/l)	(wMar) (mg/l)	Ks (hrs^{-1})	Period No.
Jan.1-31	22	12.5	2500	923	7550	2780	.01044	4.9	14	19.4	29.02	1.67	A
Feb. 1-12	25	14.2	3080	1134	8470	3120	.00591	5.1	10	13.9	18.44	2.58	B
Feb.13-17	16	8.8	3340	1232	9230	3420	.00739	4.1	13	18.1	25.30	2.24	C
Mar.12-17	14	8.1	2560	949	7780	2870	.02086	4.0	7	9.7	59.87	4.28	D
Mar.23-Apr.5	23	13.1	1850	683	7060	2610	.00694	3.0	8	11.1	18.17	2.54	E
Apr. 6-Apr.22	28	15.9	2580	943	9530	3520	.00548	2.2	10	13.9	19.29	2.92	F
Apr.23-May 3	22	12.5	3310	1220	12750	4710	.00520	1.1	14	19.4	24.49	3.37	G
May 11-May 21	25	14.2	2840	1050	12890	4760	.00462	0.9	15	20.8	21.99	3.25	H

ACTIVE MASS BALANCE
Operating Data

June - September 1971

Table 10

Period	FE-VSS (mg/l)	ML VSS (mg/l)	Ma (mg/l)	RS VSS (mg/l)	Mar (mg/l)	w	t (hrs)	FE-BOD$_5$ (mg/l)	F (mg/l)	wMar (mg/l)	Ks (hrs^{-1})	Period No.	Mae (mg/l)
[illegible]	17	1710	[illegible]	7520	2700	.00832	3.1	15	22	23.27	1.23	9	10
[illegible]	18	1100	404	5430	1280	.01342	3.3	10	19	17.18	1.99	10	10
[illegible]	10	1270	459	4530	1310	.01293	3.2	7	10	20.[illegible]	2.03	11	8
[illegible]	21	1320	577	3330	1410	.01477	3.1	5	9	20.82	2.14	12	12
8/[illegible]-8/22	23	1380	509	5150	1910	.01310	3.0	9	13	28.84	2.09	13	15
[illegible]	19	1330	573	6230	2520	.00888	3.2	12	8	20.36	3.10	14	11
[illegible]-9/12	20	1990	735	3820	3520	.00783	3.1	7	10	27.42	3.13	15	11

used as the established synthesis constant

On the basis of the above procedure, the average K_s was determined to be 2.6 hrs^{-1} for the constant detention time periods.

The following then is a summary of the various constants that were established for the CM system which are then used for further evaluations. The ultimate BOD to 5 day BOD ratio was determined from a series of rate constant tests. A better approximation was later determined to be 1.46 rather than 1.4, but 1.4 was used in the preceding calculations.

$K_s = 2.6\ hrs^{-1}$ (O_2 Units)

$K_e = .026\ hrs^{-1}$ (½ Units)

$BOD_L = 1.4 \times BOD_5$

$M_a/Vss = Mar/Vss = 0.26$

$M_{ae}/Vss = 0.40$

O_2 Uptake Tests on Plug Flow System

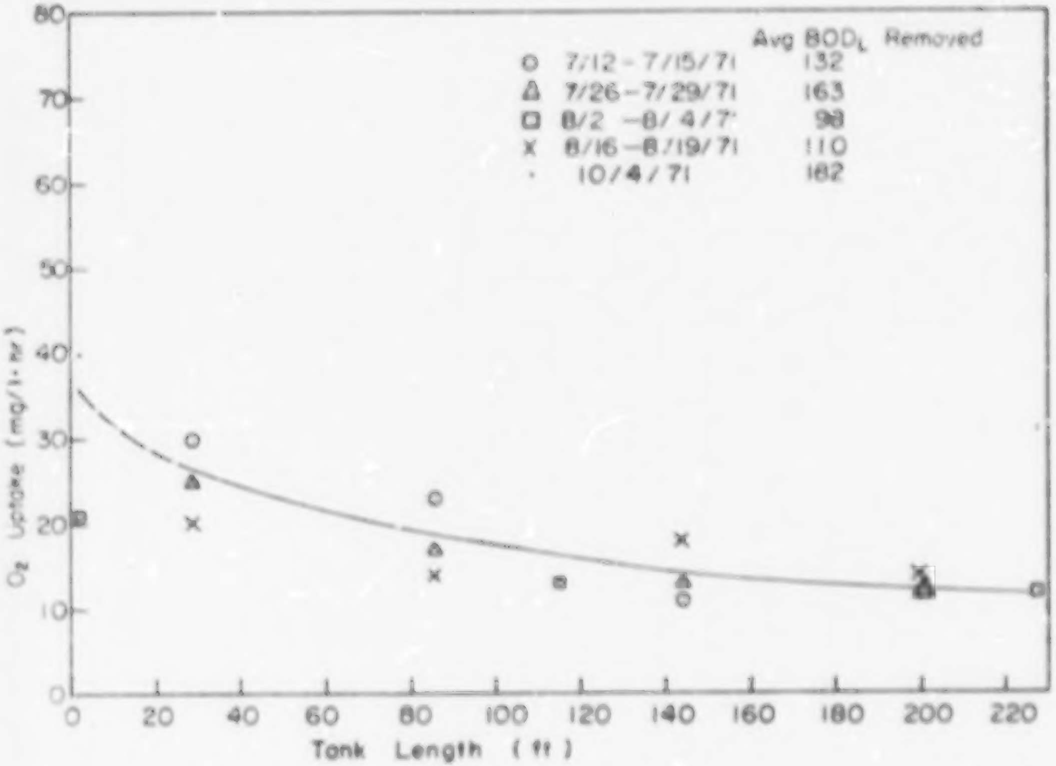
To compare oxygen requirements between the two systems a series of O_2 uptake tests were also run on the plug-flow side. Since conditions varied along the system, it was necessary to sample at a number of points to adequately describe the situation. Table 11 presents the data obtained. A key to the sample locations is shown at the bottom of Table 11. Figure 36 shows the relation of the uptake rates to the sample locations along the length of the baffled PF system.

Two basic things should affect the O_2 demand at any point in the tanks; BOD concentration and solids concentration. The average MLVSS for all of the runs shown in Table 11 (except 10/4/71) was 1150 mg/l, so it can be said that solids were relatively constant. The average ultimate BOD removed for each period is shown in Figure 36 ($BOD_L = 1.46 \times BOD_5$).

The actual uptake rate should always be something less than the total ultimate BOD removed as indicated by equation (3),

$$(F_O - F) - (1+r)\,\Delta M_a = t\,(K_s M_a + K_e F),$$

developed in Section V. The measured data bears this out. For example, take the period 7/12 - 7/15:



PLUG FLOW O_2 UPTAKE DATA

Figure 36

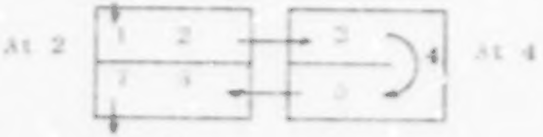
Summary of O_2 Uptake Data

Plug-Flow

Table 11

Date	Sample Location	Ks (hrs -1)	Ke (hrs-1)	F (mg/l)	Sa (O_2 Units)	Sa (solids units)	VSS (mg/l)	Sa/VSS	Total Uptake @ t = 0 (mg/l)/hr
7/12/71	2	0.83	.0311	21.7	384	272	1380	0.20	30
7/13/71	3	1.61	.0810	5.0	185	130	1200	0.11	23
7/14/71	5	1.39	.0438	2.2	183	129	710	0.18	11
7/15/71	6	1.03	.0252	3.4	357	252	640	0.39	12
7/26/71	2	1.03	.0261	14.6	383	270	1250	0.22	25
7/27/71	3	0.83	.0228	7.7	452	318	910	0.35	17
7/28/71	5	1.03	.0265	3.9	340	239	1170	0.20	13
7/29/71	6	1.03	.0236	4.5	314	221	870	0.25	12
8/2.71	1	0.83	.0223	13.2	448	316	1180	0.27	21
8/3/71	4	1.48	.0491	2.2	200	141	1050	0.13	13
8/4/71	7	0.69	.0315	[illegible]	143	101	670	0.15	12
8/16/71	2	0.72	.0308	11.0	318	224	1210	0.18	20
8/17/71	3	1.36	[illegible]	3.1	221	155	1080	0.14	14
8/18/71	5	1.66	[illegible]	4.2	309	218	1290		18
8/19/71	6	0.59	.0344	6.7	305	215	1370	0.15	14
10/4/71	1	0.59	.0393	34.0	509	358	2400	0.15	40
10/ 4/71	7	1.03	.0298	14.0	534	330	2410	0.16	31

Key to Sample Locations



$\frac{F_o - F}{t} = \frac{132\ mg/l \times 1.71\ MGD}{.264\ MG \times 24\ hr/day} = 35.5\ \frac{mg/l}{hr}$ as compared with a maximum measured rate for that period of 30 $\frac{mg/l}{hr}$.

Air Flow versus D.O. Tests

To obtain some information about O_2 demand from actual full scale operating conditions, probes were placed in the aeration tanks for 24 hour periods and an attempt was made to hold the DO constant by varying the air flow.

Figures 37 and 38 show the results of one such test for CM and PF. It was not possible to hold the DO constant, but the average DO for each side correlated quite closely (3.4 mg/l for CM and 3.6 mg/l for PF). During the test equal sewage flow of equal strength was applied to each side. Average mixed liquor temperature was 21°C.

To compare the two sides it is necessary to know the transfer efficiency of each system. The next section discusses the tests performed to determine this. The transfer efficiency data has 2 limitations. First, only 1 test was performed on each side and this was at approximately 500 scfm. As air flow increases, efficiency normally decreases. Second, due to slow deoxygenation during the PF test, that result is questionable.

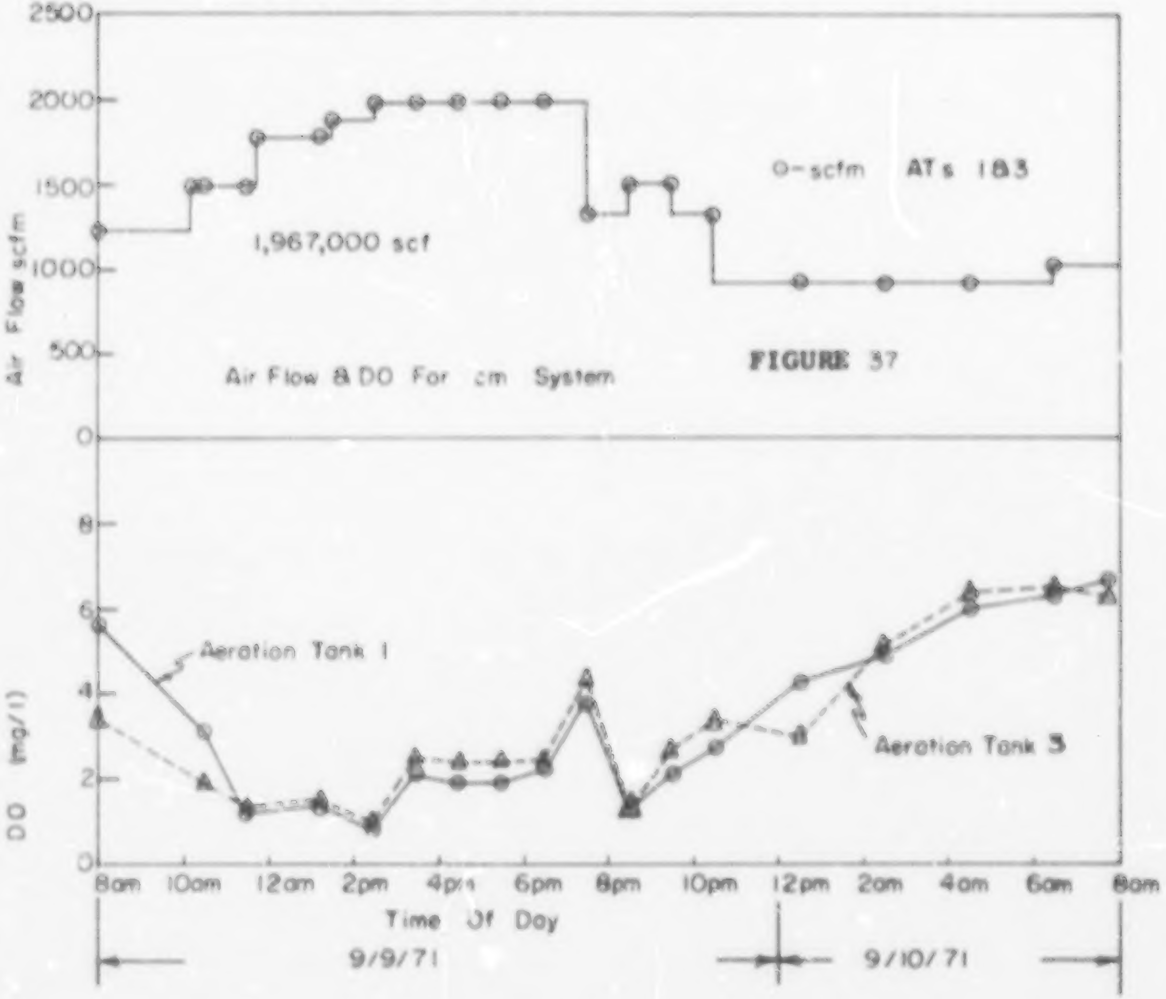
Some analysis, however, can be made. Four areas should be considered:

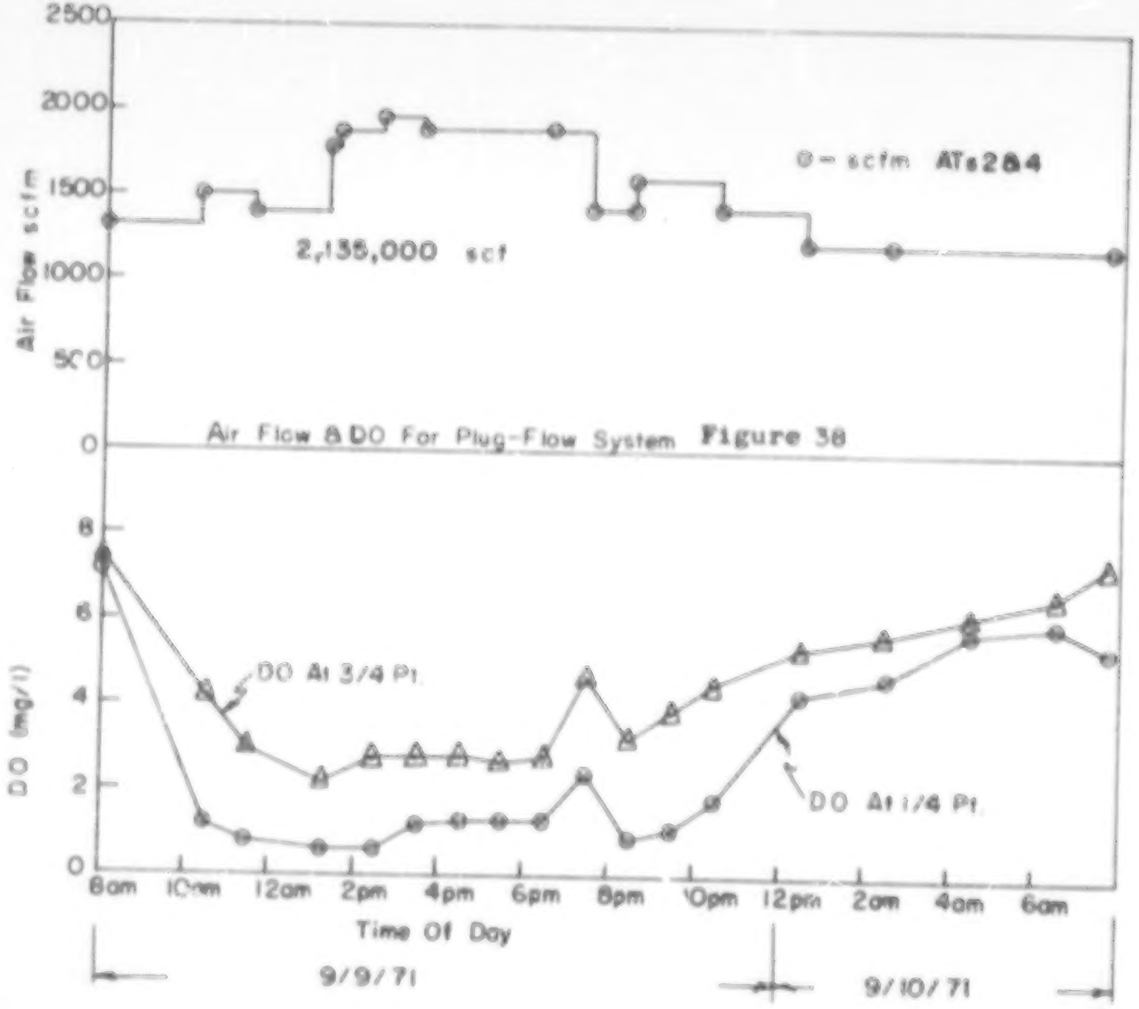
1. Air applied and transfer efficiency
2. Ultimate BOD removal
3. Solids or active mass concentration
4. Application of established reaction rate constants to predicting O_2 required.

Referring to Figure 37, 1,967,000 scf of air was applied from 8 am 9/9/71 to 8 am 9/10/71. Assuming 10% efficiency (determined for CM at 500 scfm, the following calculations can be made:

$$1,967,000\ \frac{scf}{day} \times .10 \times \frac{.0174\ lb\ O_2}{scf} \times \frac{1}{.264\ mg \times 8.33\ \frac{lb}{gal} \times 24\ \frac{hrs}{day}}$$

$$= 64.8\ \frac{mg/l}{hr}\ (@\ 20^{o}C)$$





Air Flow & DO For Plug-Flow System **Figure 38**

BOD removal was the same for both systems on this particular day and the average MLVSS compared very closely (2140 mg/l for CM and 2020 mg/l for PF).

BOD_5 values: $\frac{\text{unfilt. PE}}{88\text{ mg/l}}$ $\frac{\text{filt. FE,CM}}{10}$ $\frac{\text{filt. FE,PF}}{10}$ $\frac{\text{removed}}{78}$

The maximum O_2 rate based on the ultimate BOD removed is equivalent to the following for each system:

$$.5 \times \frac{4.615\text{ MGD} \times 78\text{ mg/l} \times 1.46 \times 8.33\ \frac{\text{lb}}{\text{gal}}}{.264\text{MG} \times 833\ \frac{\text{lb}}{\text{gal}} \times 24\text{ hr/day}} = 41.5\ \frac{\text{mg/l}}{\text{hr}}$$

The following equation is applicable to the CM system:

$$\frac{do}{dt} = 2.6F + .037\text{ Ma},\ \text{Ma} = \text{MLVSS} \times .26 = 2140 \times .26 = 556$$

$$\frac{do}{dt} = 2.6 \times 10 + .037 \times 556 = 46.6\ \frac{\text{mg/l}}{\text{hr}} \quad (@\ 20^\circ C)$$

This indicates that the rate calculated from the transfer efficiency of 10% is probably a little high. An efficiency of 7% would yield 45.4 $\frac{\text{mg/l}}{\text{hr}}$. An efficiency of 8% would give 51.8 $\frac{\text{mg/l}}{\text{hr}}$.

Now referring to Figure 38, the total air applied to the PF system for 24 hours was 2,135,000 scf. The transfer efficiency measured in the field for PF was 3.5%. This would yield a rate of 24.6 $\frac{\text{mg/l}}{\text{hr}}$. If you assume air requirements are equal in each system, then an estimate of the transfer efficiency in the PF can be made by taking the ratio of the air applied to each times the efficiency in the CM:

$$\frac{1,967,000}{2,135,000} \times 8\% = 7.3\% \text{ or } .921 \times 7\% = 6.4\%$$

O_2 Transfer Efficiency Tests

A series of tests were conducted on the full scale and pilot plant aeration tanks using the sodium sulfite reoxygenation method of measuring O_2 transfer.[12, 13] The results are shown in Table 12.

α values were determined from lab scale tests, and β values were obtained by aerating the various test liquids to saturation, i.e., constant DO.

Summary of O_2 Transfer Tests

Table 12

Test Conditions	Air Applied scfm	$\frac{scfm}{Diff.}$	$\frac{scfm}{1000 ft^3}$	K_{La} @ 20°C (min^{-1})	Trans. Eff. @ 20° C.	Projected Eff. in Mixed Liquor
Full Scale						
a) AT 1, 748	532	0.71	30.2	.0823	10.9%	12.8%*
Diffusers						
Pri. Effluent						
b) At 1,242 Dif.	531	2.19	30.2	.0701	8.9	10.4%*
Prim. Effluent						
c) AT 2 156 Dif.	500	3.21	28.3	.0224	3.0	3.5*
Pri. Effluent						
d) AT 4,102 Dif.	480	4.71	27.2	.0190	2.7	3.2*
Prim. Effl.						
Pilot Plant						
a) Tap Water	.051	.051	15.3	.0803	18.8	10.7†
b) Tap Water	.136	.136	40.7	.202	17.8	10.1†
c) Tap Water	.260	.260	77.8	.314	14.6	8.3†
d) Final Eff.	.060	.060	18.0	.0638	13.0	13.0
e) Final Eff.	.136	.136	40.7	.119	10.3	10.3
f) Final Eff.	.260	.260	77.8	.167	7.6	7.6
Lab. Scale						
a) 1 liter beak.	.00565	.00565	160	.0423	1.19	-
diffuser stone						
Final Eff.						
b) Same as a)	.00565	.00565	160	.0360	1.01	-
But Pri. Eff.						

* based on an $\alpha = \frac{K_{Fe}}{K_{Pe}} \cong 1.17$

† based on an average $\alpha = \frac{K_{Fe}}{K_{H_2O}} = 0.57$

Only one major problem was experienced, and this occurred in aeration tanks 2 & 4. Dispersion of the sodium sulfite was slow, and it appeared that insufficient cobalt catalyst was present because it took over 30 minutes for the DO to drop to the 1-2 mg/l range. It did not go all the way to zero. Probably what happened during the re-oxygenation was continued reaction of some sulfite tending to depress the DO and reduce the rate of DO increase yielding a lower measured transfer rate.

Parallel O_2 Uptake Tests

Finally an attempt was made to compare measured O_2 uptakes from both systems at the same point in time under similar loading conditions. The method already described was used to measure O_2 uptake rate except that each run in the sealed flask was only continued for about 1 hour instead of the usual 24.

During a period from 11 am to 4 pm on November 16, 1971, the following procedures were carried out:

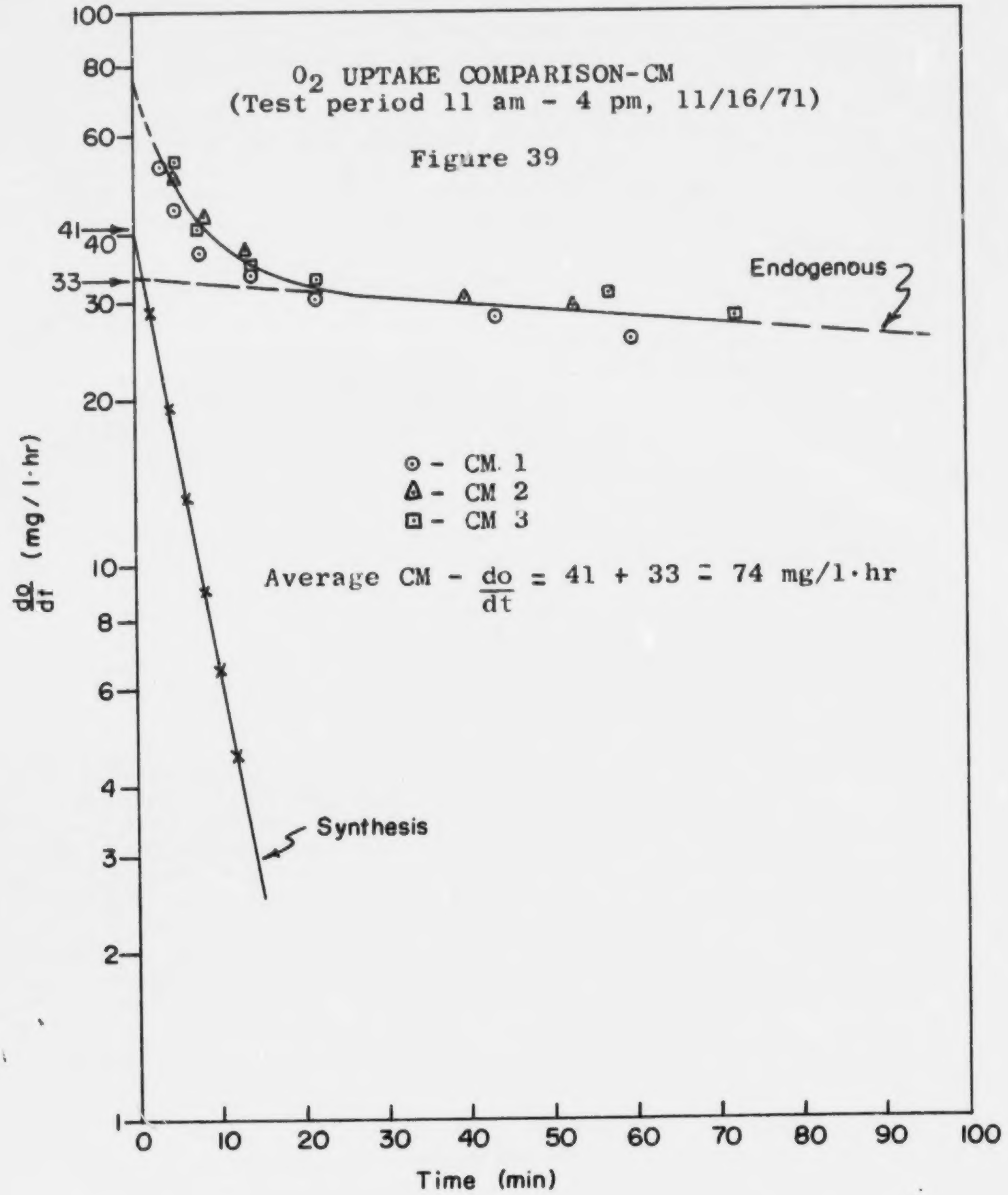
1. A sample was drawn from the east end of aeration tank 1 (CM) at about 11 am, 1 pm and 2:30 pm.
2. At the same time another sample was drawn from the inlet area, the half way point, and the outlet area of aeration tanks 2 & 4, the PF system.

Figure 39 shows the results of the three tests conducted on the CM system. The procedure (outlined in detail in the appendix) for determining the initial oxygen uptake rate consisted of determining the slope of the endogenous portion (flat portion) of the curve and subtracting this from the entire curve to obtain the slope of the synthesis curve. The two resulting straight lines were then extrapolated back to t = 0 to find the respective contributions to oxygen demand (K_eM_a and K_sF) existing in the mixed liquor sample. This analysis yielded an oxygen uptake rate of 74 mg/l·hr.

This can be compared with the predicted rate using the equation $\frac{do}{dt} = K_sF + K_eM_a$.

If this equation is applied to this situation, it would predict the following for the CM system:

Available Data
Average MLVSS = 3600 mg/l
Average FE BOD_L = 19 mg/l (average of November 14, 15,17, and 18; no data available for November 16)



O_2 UPTAKE COMPARISON-CM
(Test period 11 am - 4 pm, 11/16/71)

Figure 39

Average CM - $\frac{do}{dt}$ = 41 + 33 = 74 mg/l·hr

$$M_a = .26 \times MLVSS = .26 \times 3600 = 936$$

Then:

$$\frac{do}{dt} = 2.6 \times 19 + .037 \times 936 = 19 + 35 = 84 \text{ mg/l·hr}$$

This value of 84 mg/l·hr is in reasonably good agreement with the measured value of 74 mg/l·hr.

It is of interest from a design standpoint to relate the above uptake rate to the conventional design parameter scf air/lb BOD_5 removed. This can be calculated as follows: assuming an uptake rate of 74 mg/l·hr

$$\frac{74 \text{ mg}}{\text{l·hr}} \times .264 \text{ MG} \times 8.33 \frac{\text{lb}}{\text{gal}} \times \frac{24 \text{ hr}}{\text{day}} = 3910 \text{ lb } O_2\text{/day}$$

$$\frac{3910 \text{ lb/day}}{.08 \times .0174 \text{ lb } O_2\text{/scf}} = 2,809,000 \text{ scf/day}$$

$$\frac{2,809,000 \text{ scf/day}}{127 \frac{\text{mg}}{\text{l}} BOD_5 \text{ removed} \times 2.5 \text{ MGD} \times 8.33} = 1060 \text{ scf air/lb}$$

BOD_5 removed. This value of approximately 1000 scf/lb will be used in the design analysis section of the report.

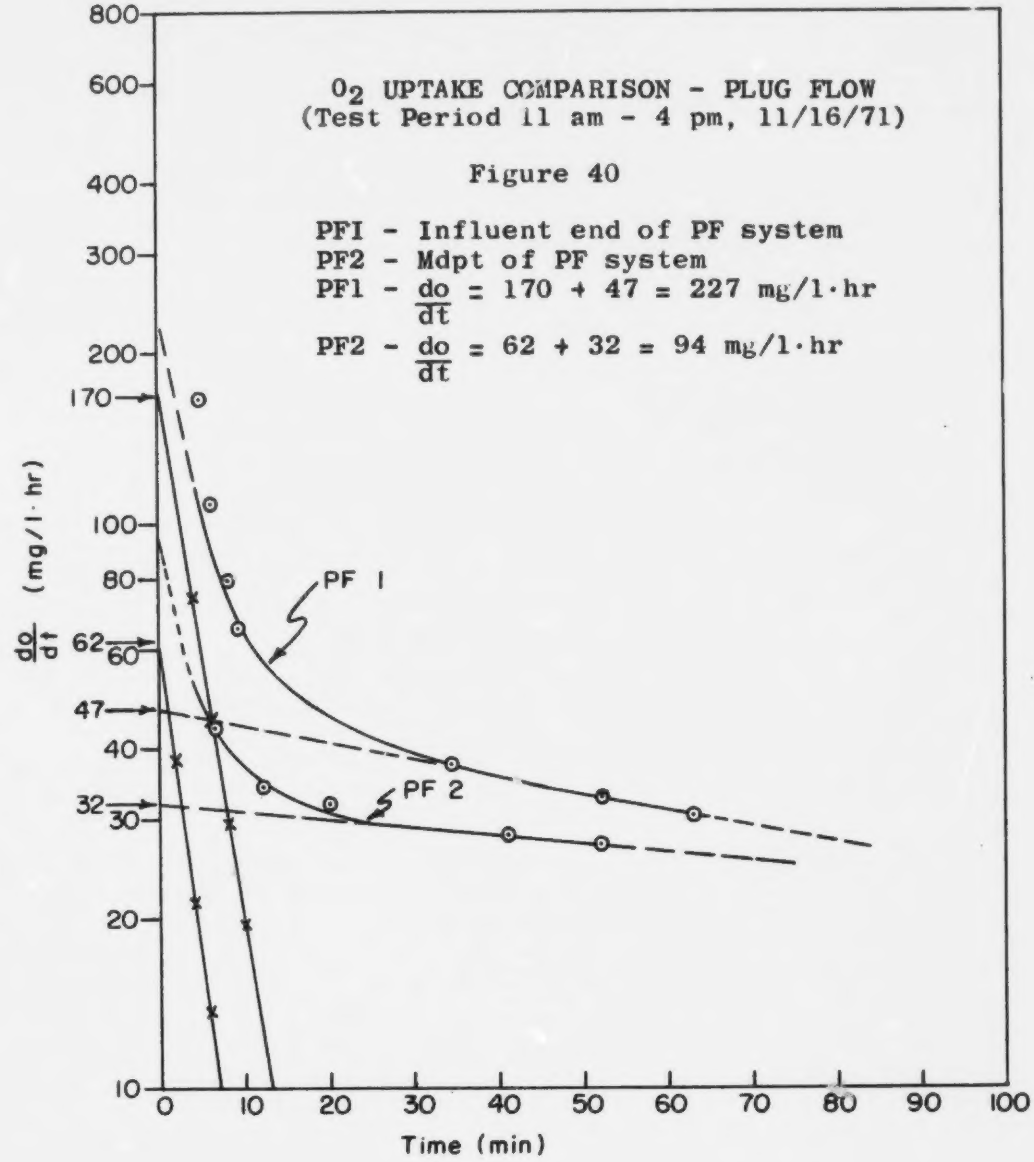
Figures 40 and 41 show for PF the uptake analysis described previously. Uptake rates ranged from 227 mg/l·hr at the head end of the PF system to 48 mg/l·hr at the effluent. These values are shown in a plot uptake versus tank length in Figure 42. The average rate was calculated by dividing the Figure 42 plot into 23 length increments, finding the $\frac{do}{dt}$ at the midpoint of each increment and figuring the average of these rates. This yielded a value of 107 mg/l·hr. This value could change appreciably depending on what occurs between the three points sampled.

Solids Production

Solids wasting data was, of course, collected throughout the period of the study. For the purpose of the analysis, data was considered for the period January 1, 1971 to October 1, 1971. Data covers both constant and variable flow conditions.

One approach to developing a prediction equation for solids production involved use of the active mass balance. This utilizes the equation already developed in Section V:

$$M_{ae} + wM_{ar} = T(K_sF - K_eM_a)$$



O_2 UPTAKE COMPARISON - PLUG FLOW
(Test Period 11 am - 4 pm, 11/16/71)

Figure 40

PFI - Influent end of PF system
PF2 - Mdpt of PF system
PF1 - $\frac{do}{dt} = 170 + 47 = 227$ mg/l·hr
PF2 - $\frac{do}{dt} = 62 + 32 = 94$ mg/l·hr

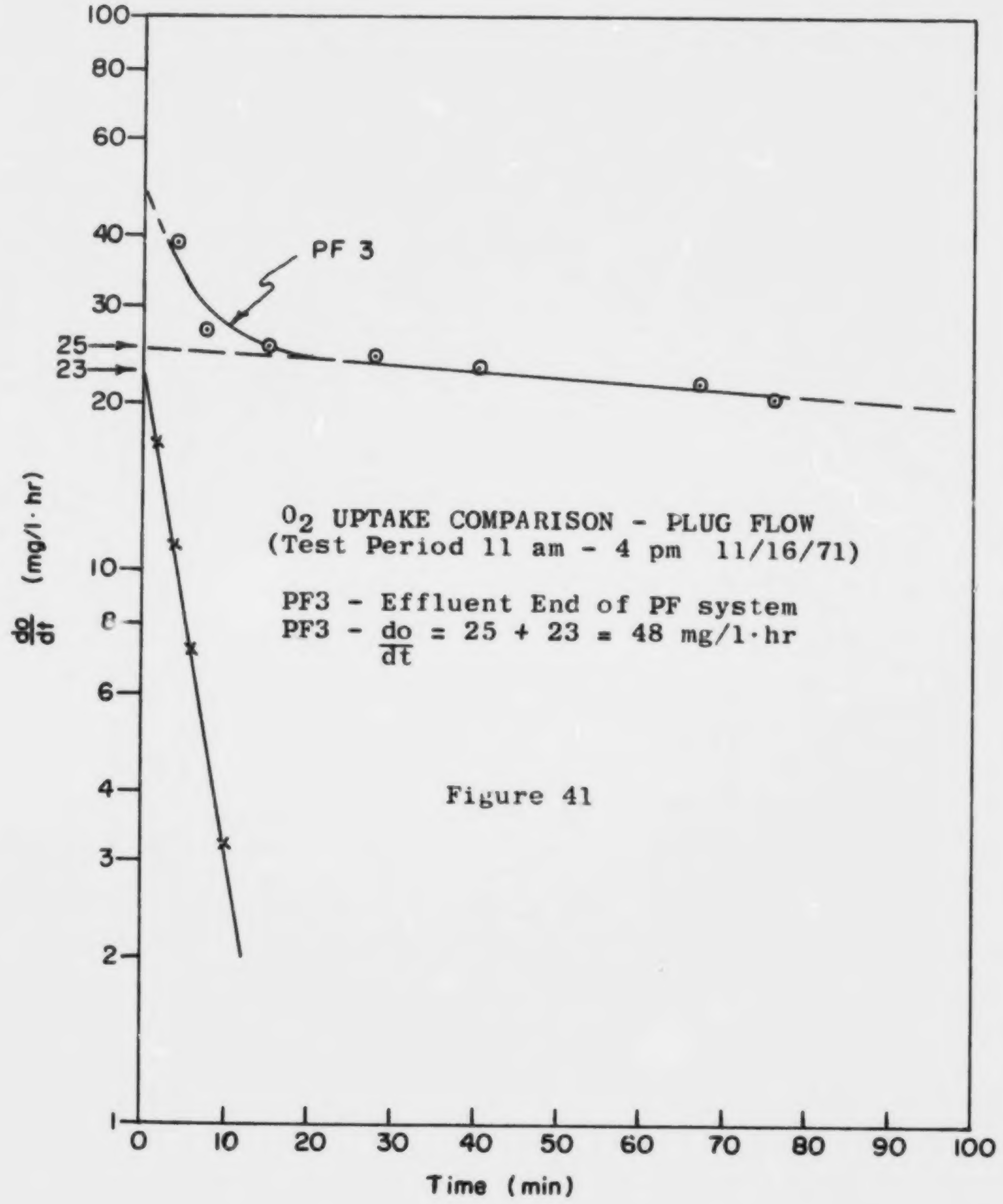
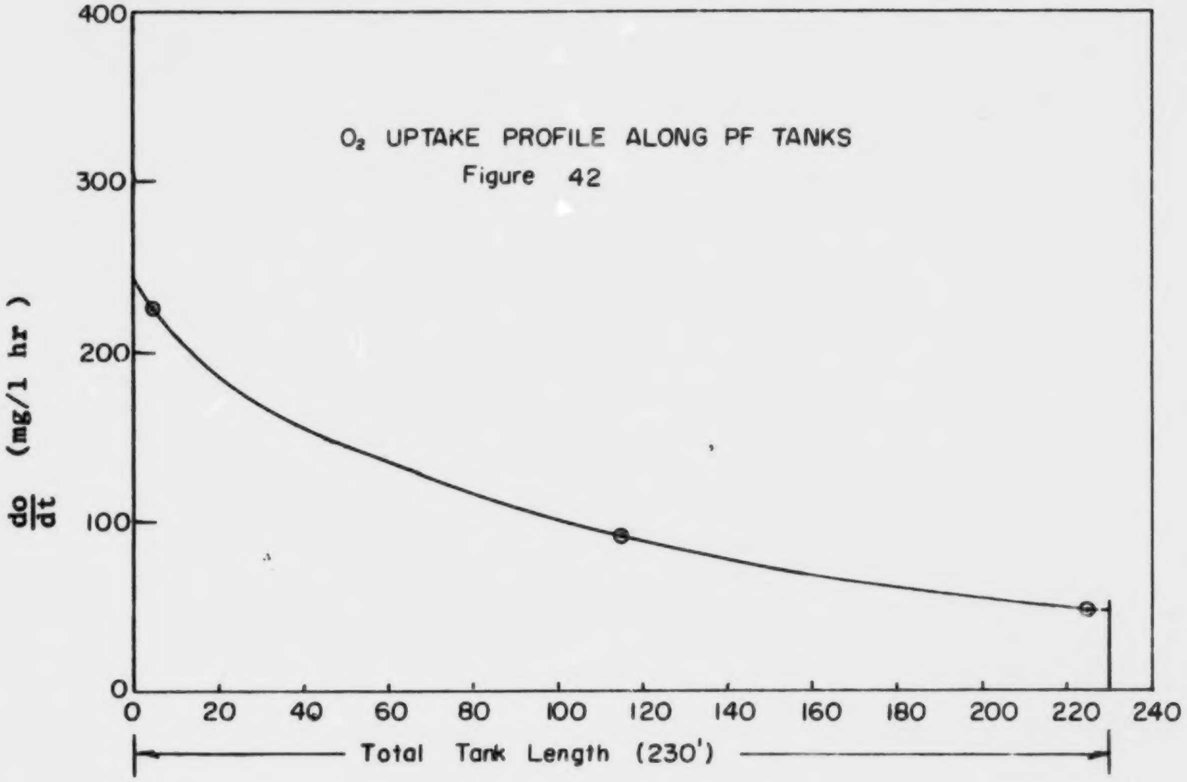

O_2 UPTAKE COMPARISON - PLUG FLOW
(Test Period 11 am - 4 pm 11/16/71)
PF3 - Effluent End of PF system
PF3 - $\frac{do}{dt}$ = 25 + 23 = 48 mg/l·hr
PF 3
25
23
100
80
60
40
30
20
10
8
6
4
3
2
1
0
10
20
30
40
50
60
70
80
90
100
Time (min)
$\frac{do}{dt}$ (mg/l·hr)

Figure 41



O_2 UPTAKE PROFILE ALONG PF TANKS
Figure 42

To evaluate this approach the following calculations were made:

First, by considering the equation,

$M_{ae} + wM_{ar} = T(K_sF - K_eM_a)$, and using the constants established from Tables 7 - 10, it is possible to solve for the fraction of incoming flow wasted, w:

$$w = \frac{T(K_sF - K_eM_a) - M_{ae}}{M_{ar}}$$

The mass of VSS wasted per day (ΔX) is then given by:

$$\Delta X = \frac{(QwM_{ar})\ 8.33\ \text{lb/gal}}{0.26 \times 1.42} = QwM_{ar} \times 22.6\ \frac{\text{lb VSS}}{\text{day}}$$

where Q = influent flow in MGD
M_{ar} = active mass in return sludge stream (mg/l, O_2 units)

If you define ΔM_a as the net change in active mass concentration of the entire flow (Q + rQ) as it moves through the aeration tank, the following equation can be written:

$$\Delta M_a\ (Q + rQ) = K_sFV_a - K_eM_aV$$

In other words the left hand term in the above equation is equivalent to the active mass produced and can be related to the mass of VSS produced by the M_a/VSS ratio of 0.26.

The equation can be rewritten:

$$\Delta M_a\ (1 + r) = T(K_sF - K_eM_a)$$

Since the right hand term has already been calculated and the recirculation ratio r is known for each period, ΔM_a can be calculated from:

$$\Delta M_a = \frac{T(K_sF - K_eM_a)}{(1 + r)}$$

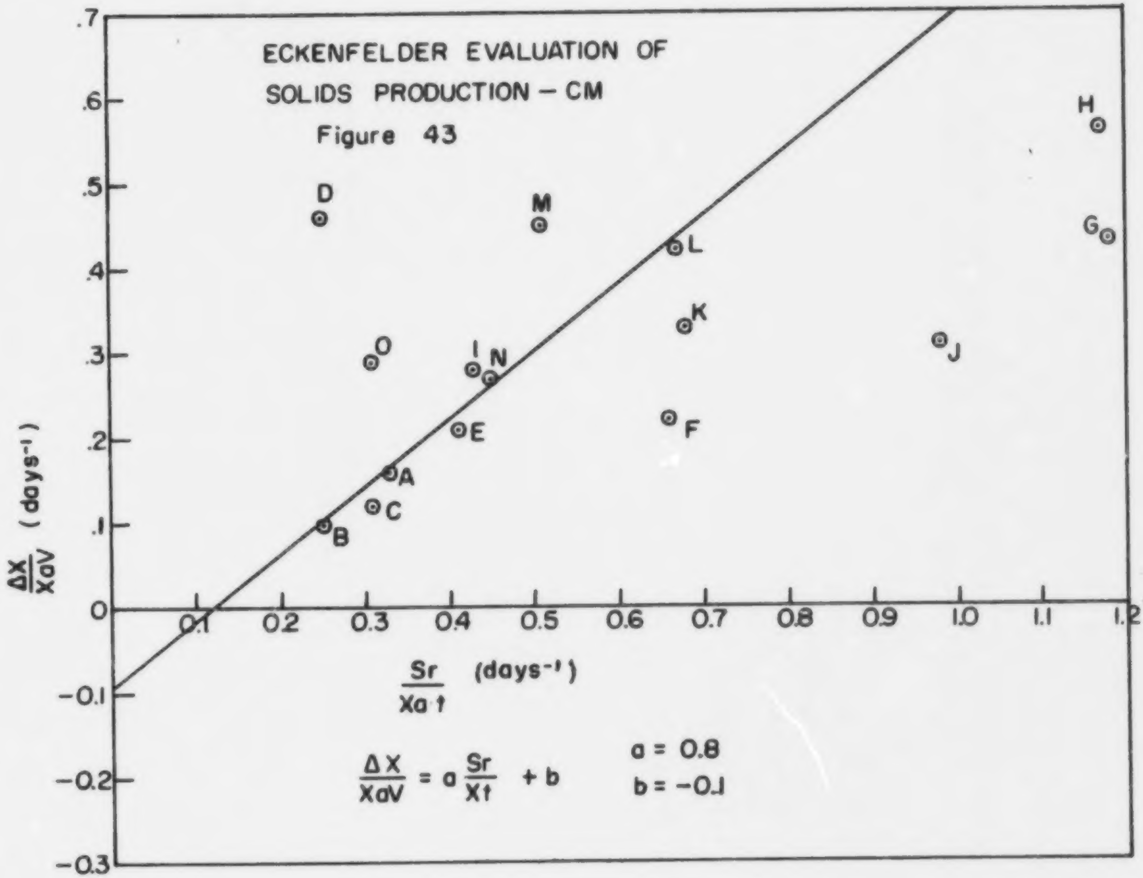
If $\Delta X'$ is defined as the mass of VSS produced per day, then it can be calculated as follows:

$$\Delta X' = \overbrace{(Q + rQ)}^{\text{MGD}} \times \overbrace{\Delta M_a}^{\text{mg/l, } O_2 \text{ units}} \times \frac{1 \text{ solid unit}}{1.42\ O_2 \text{ units}} \times$$

Period	Pounds of Volatile Suspended Solids per Day			
	Active Mass Balance		Eckenfelder Approach	Measured
	ΔX	$\Delta X'$	ΔX	ΔX
A	3460	3819	940	854
B	546	944	680	690
C	2980	2140	1100	915
D	-	88	560	2583
E	950	1562	900	853
F	757	1779	2420	1239
G	522	1326	3060	1579
H	790	1919	2590	1756
I	5340	5803	940	1067
J	5174	5614	1650	750
K	1725	1998	1230	944
L	1390	1948	990	961
M	2290	2903	940	1351
N	340	832	890	928
O	474	979	660	1267

(ΔX = mass wasted per day, $\Delta X'$ = mass produced per day)
COMPARISONS OF SOLIDS PREDICTION EQUATIONS FOR COMPLETE MIX OPERATION

Table 13



ECKENFELDER EVALUATION OF SOLIDS PRODUCTION – CM

Figure 43

$$\frac{1 \text{ lb VSS}}{.26 \text{ lb } M_a} \times \frac{8.33 \text{ lb}}{\text{gal.}} = \frac{\text{lb. VSS produced}}{\text{day}}$$

Table 13 is a summary of the results of using the method of solids production prediction discussed above. Table 13 also includes the results of analyzing the data using the Eckenfelder approach outlined in Section V.

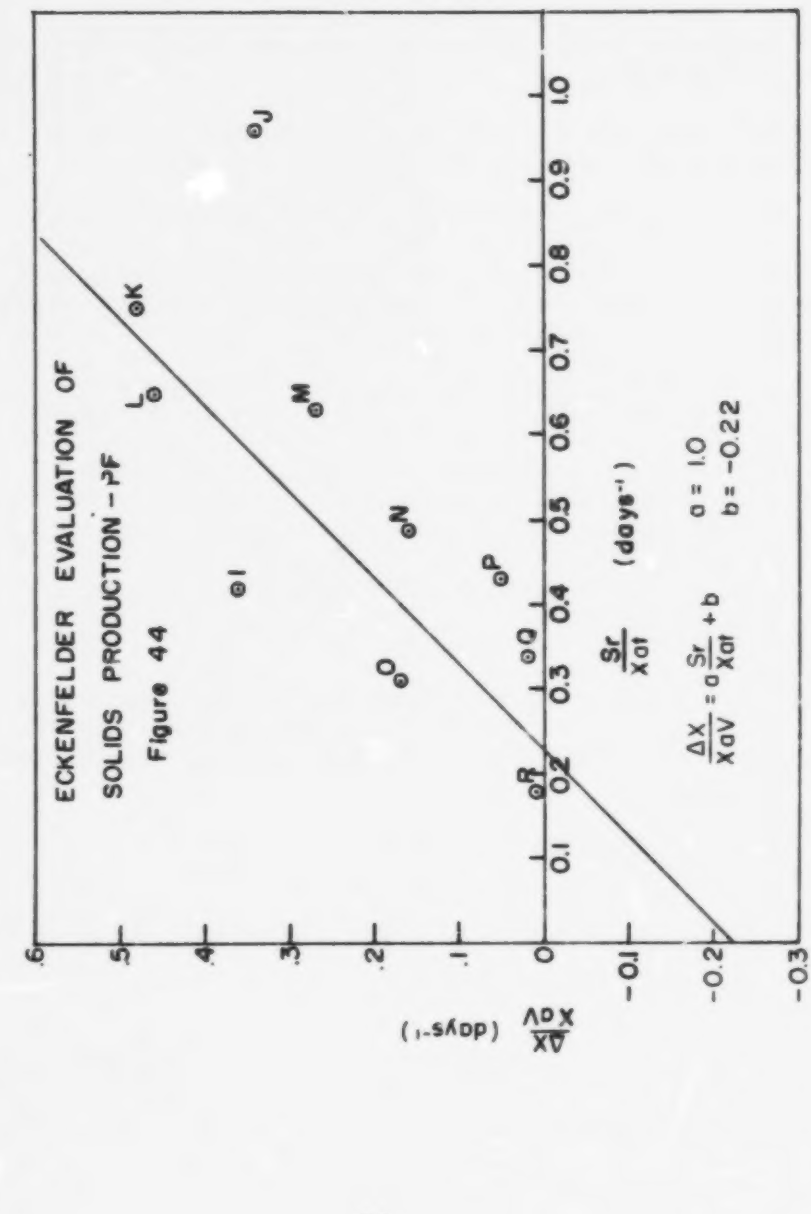
Figures 43 and 44 show plots of $\Delta X/X_aV$ versus S_r/X_at for CM and PF. The parameters are defined in detail below:

ΔX = lb VSS wasted/day
X_a = MLVSS (mg/l)
V = tank volume millions of gallons
X_aV = lbs of MLVSS = MLVSS (mg/l) x V(MG) x 8.33(lb/gal)
Sr = ultimate BOD removed = PE unfiltered BOD_L - FE unfiltered BOD_L
t = detention time (days) = Q/V

The lines drawn through the points in Figures 43 and 44 are merely to indicate the trend. They are not the result of statistical analysis.

The predicted values of solids wasted under "Eckenfelder Approach" in Table 13 were determined by using the S_r/X_at value for each period to find the corresponding $\Delta X/X_aV$ from the trend line in Figure 43. Then by multiplying this by the known X_aV the ΔX is calculated and reported in Table 13.

Both methods of predicting solids production show considerable scatter. This is one area that still needs further investigation. Better mathematical models are required that take into account the fate of all influent solids and consider solids lost in the final settling tank effluent. Also, improvements are needed in laboratory solids determination and measurement of waste sludge streams.



ECKENFELDER EVALUATION OF SOLIDS PRODUCTION - PF

Figure 44

$$\frac{\Delta X}{XaV} = a\frac{Sr}{Xat} + b$$

Final Settling Tank Performance

Though the primary purpose of the project was the evaluation of aeration tank performance under varying conditions, considerable data was also gathered concerning performance of the final settling tank (FST) system. The data arose from three basic sources:

a) The CM constant detention time phase of operation

b) Two periods during September and November 1971 when above-normal loads were applied to the settling tanks

c) A tracer test on a final tank conducted in conjunction with an aeration tank tracer test

CM Constant Detention Time Period

During this period the flow rate to the CM side utilizing two settling tanks ranged from 1.3 MGD to 3.5 MGD resulting in loadings to the clarifiers as shown below:

surface settling rate	200-700 gpd/ft^2
solids loading	10-30 lb/ft^2.day

To describe FST performance, four treatment parameters (final effluent TSS and VSS, and percent removal of TSS and BOD_5) were plotted against surface settling rate and solids loading. The data for the constant T,CM period are shown in Figures 45,46,47 and 48 as triangular points.

Above Normal FST Loadings

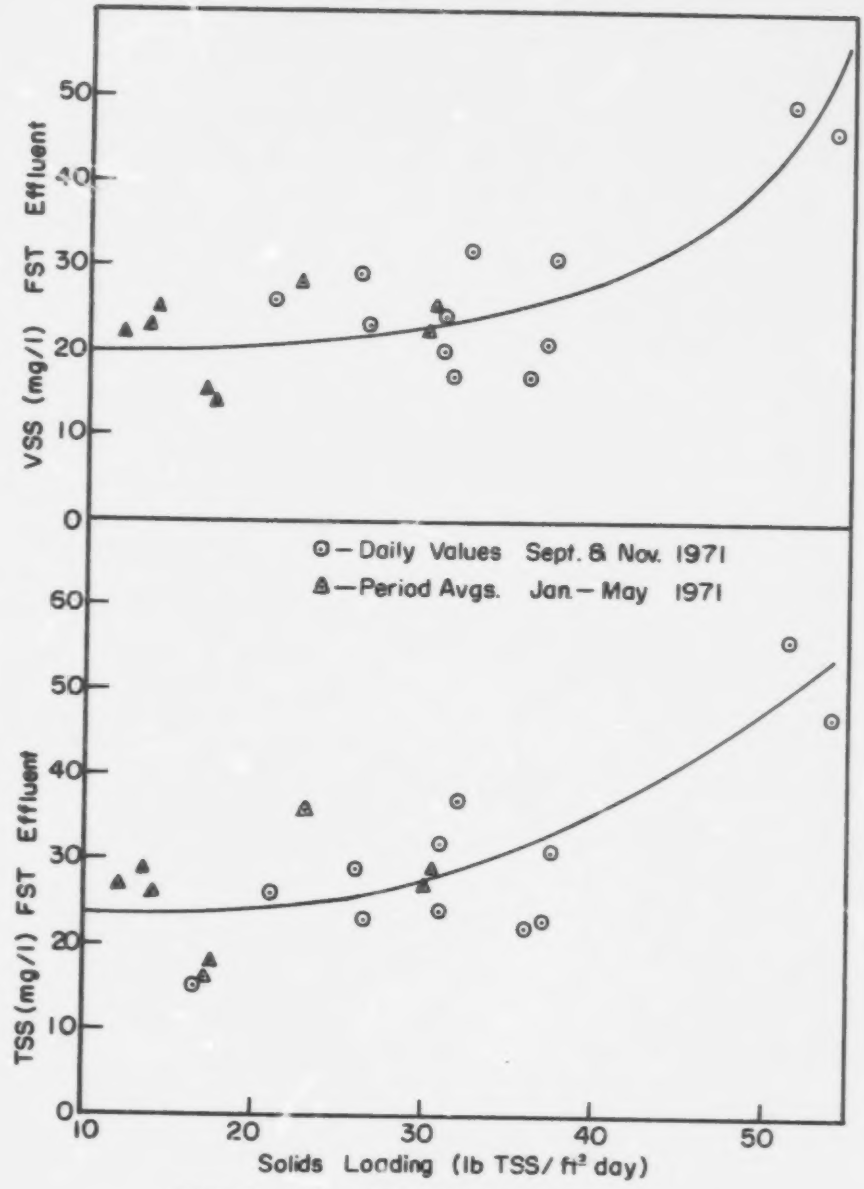
For a period of time in September and again in November 1971, flow to 2 of the 4 FSTs was cut off resulting in increased loadings on the remaining two tanks. The loading ranges are shown below:

surface settling rate	600-1400 gpd/ft^2
solids loading	20-55 lb/ft^{2}day

The data for these two periods are shown as circular points in Figures 45 through 48.

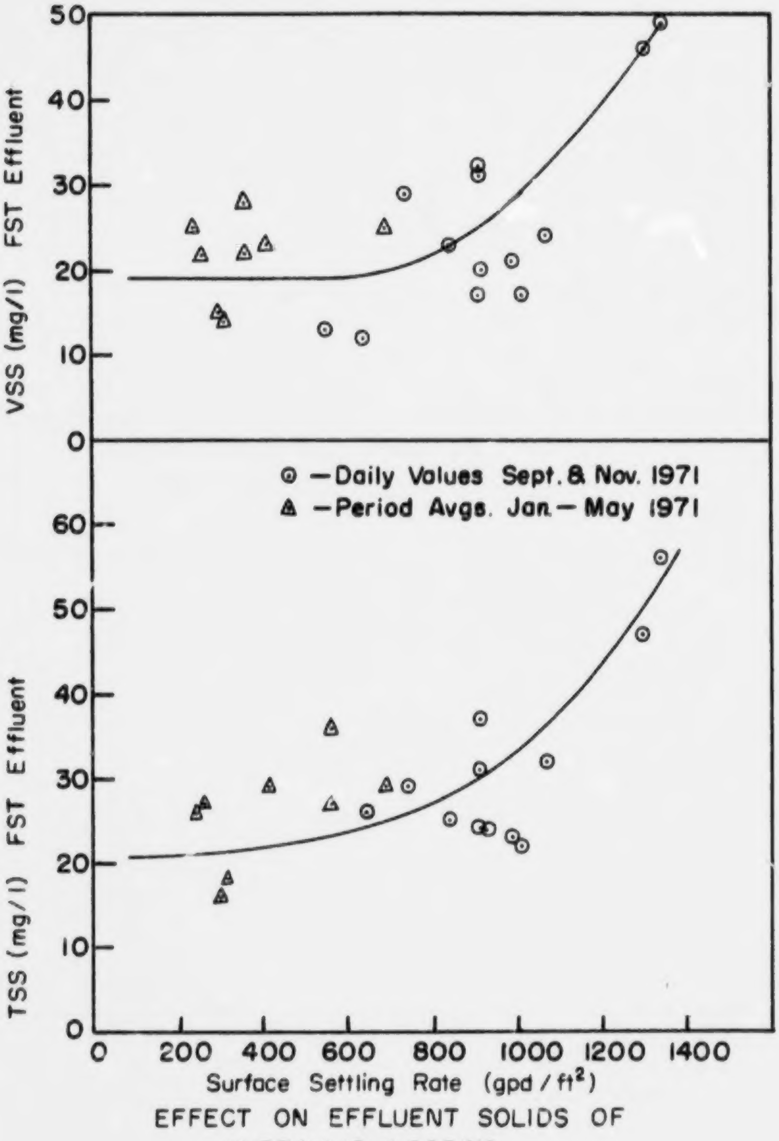
FST Tracer Test

On October 22, 1971 a tracer test was conducted using the method described in the tracer study phase of this report. An additional set of grab samples were taken from the eff-



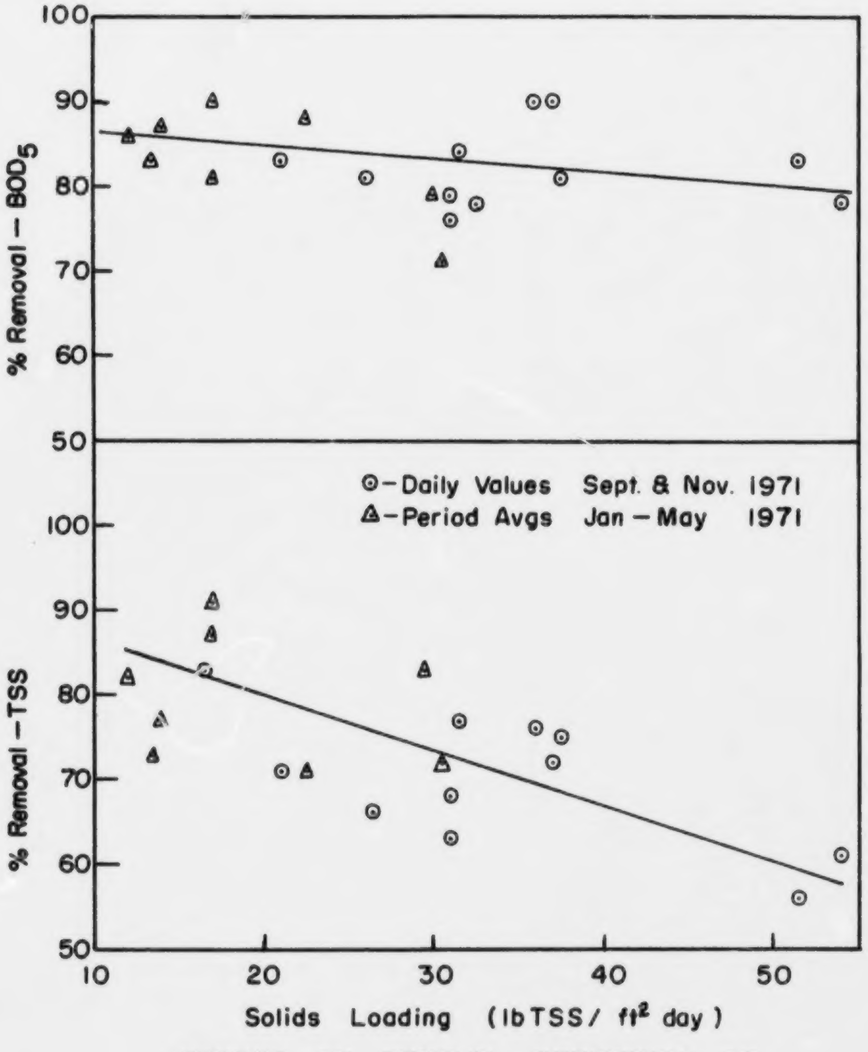
EFFECT ON EFFLUENT SOLIDS OF SOLIDS LOADING

Figure 45



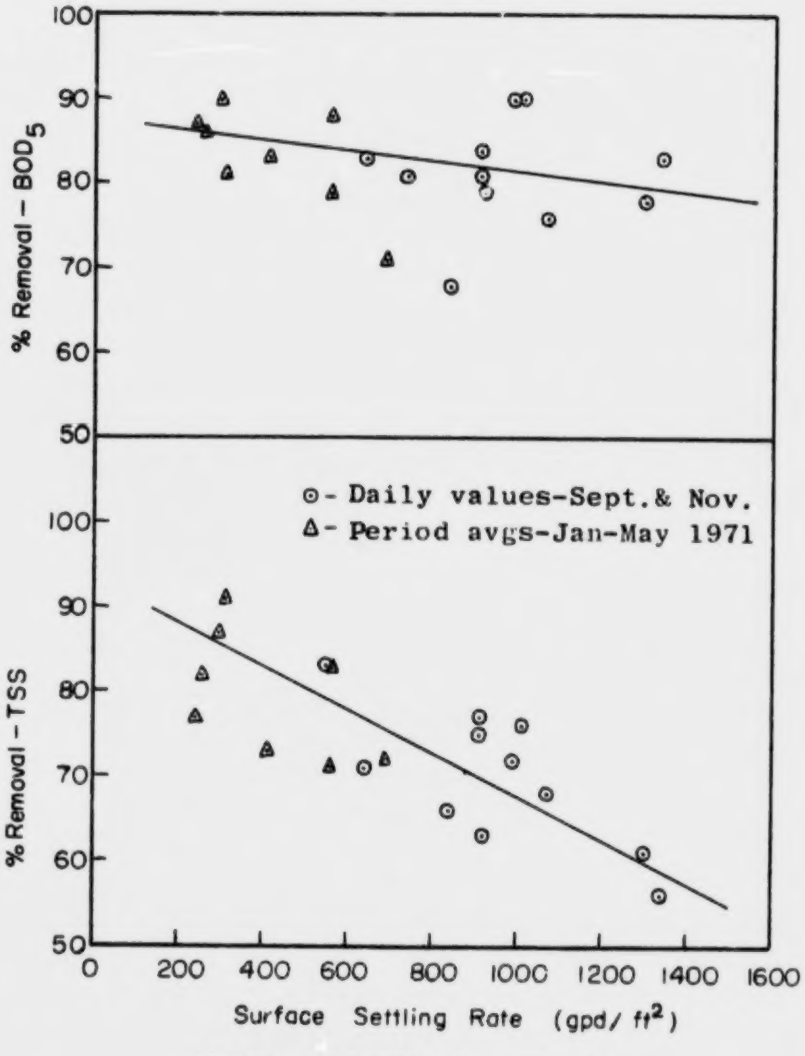
EFFECT ON EFFLUENT SOLIDS OF HYDRAULIC LOADING

Figure 46



EFFECT ON REMOVAL EFFICIENCY OF SOLIDS LOADING

Figure 47



EFFECT ON REMOVAL EFFICIENCY OF HYDRAULIC LOADING

Figure 48

luent of FST No. 1 at a point 180° opposed to the inlet channel. The samples were analyzed for dye concentration using the fluorometer and the results are shown in Figure 49.

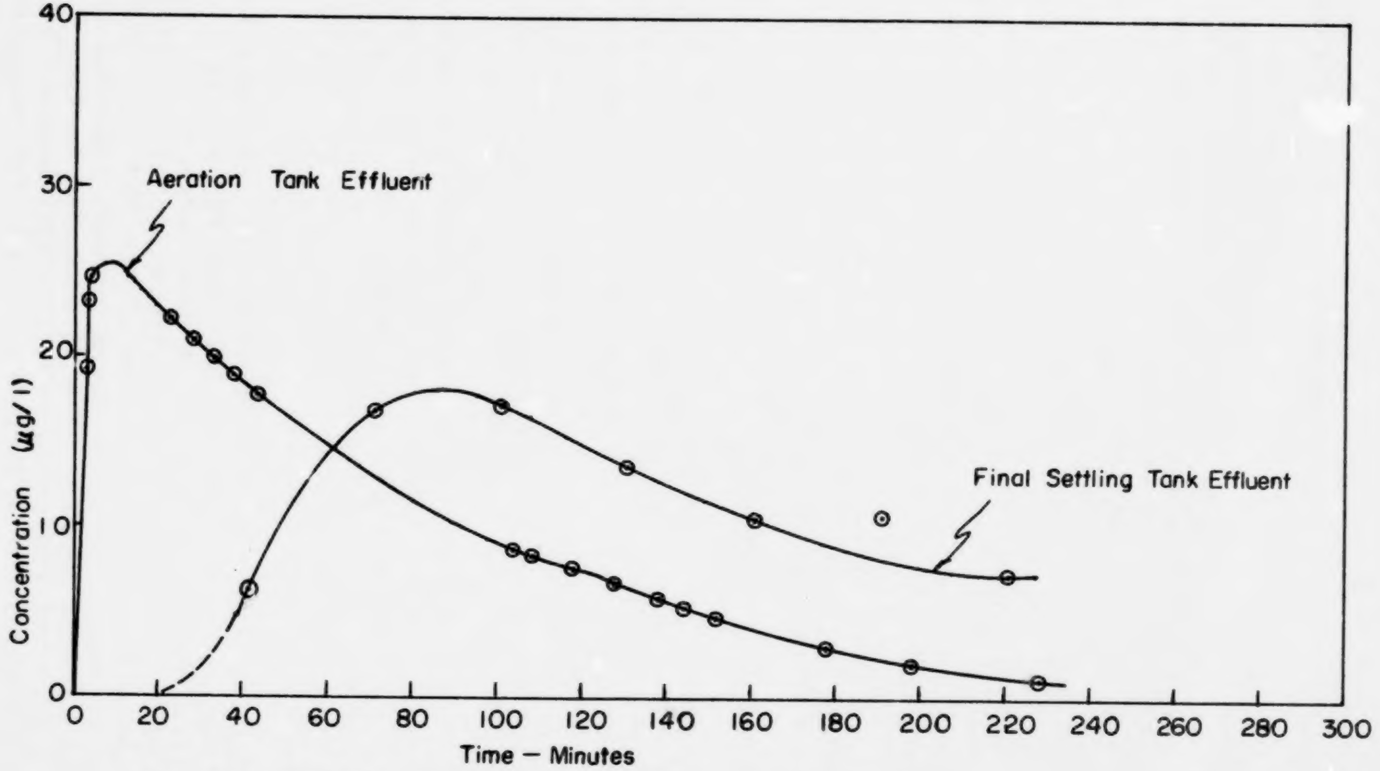
The first curve depicts the change in dye concentration in the aeration tank effluent. It also represents the input of dye to the FST. The second curve shows the change in dye concentration in the FST effluent. Since the input to the tank was not an ideal pulse, it is not possible to apply the usual tracer mathematics to this situation. However, by comparing the time between the peak concentration (inlet to outlet), it is possible to show the general timing of liquid flow through the system. The time between peaks was 79 minutes or about 1.33 hrs. (It should be emphasized at this point that the discussion is concerned only with the liquid portion of the flow and not the sludge portion). During this test the flow rate to the tank was 2.55 mgd. For a tank volume of 270,000 gallons, this yields a nominal detention time of 2.54 hours or about twice the peak-to-peak time measured above.

Discussion of FST Performance

Probably the best criterion for activated sludge FST performance is the actual solids concentration in the final effluent. Primary effluent to final effluent percent removal is not as meaningful because you are really considering removal of solids generated in the aeration tanks more than removal of solids introduced by the primary stream. Figures 45 and 46 show the effects of varying solids and hydraulic loadings on final effluent solids concentrations.

If a maximum allowable FE TSS level of 25 mg/l is chosen, these figures indicate a maximum solids load of 25 lb/ft^2 day and a maximum surface settling rate of 700 gpd/ft^2. These loadings are recommended with the realization that there is interaction between solids and hydraulic loading which may mask the effect on one or the other.

Figures 47 and 48 show the effect of solids and hydraulic loading on removal efficiency of BOD_5 and TSS across the complete activated sludge system. As has been stated, these relationships depend also on the performance of the aeration tanks and are not as representative of the FST efficiency alone. They do show a definite trend towards improved efficiency with reduced loading. The recommended surface settling rate of 700 gpd/ft^2 predicts a BOD_5 removal efficiency of 83% and a TSS efficiency of 75%. The recommended solids loading of 25 lb/ft^2 day predicts BOD_5 and TSS removal efficiencies of 84% and 77% respectively.



TRACER TESTS ON AERATION TANK & FINAL SETTLING TANK

Figure 49

SECTION VIII

SHOCK LOAD STUDIES

Introduction

Since no great differences were found in system performance under normal operating conditions, it was decided to apply some shock loads to determine the response of each side to rapid changes in loading.

To evaluate performance grab sampling was done. During each test samples were taken periodically at the aeration tank influent and final settling tank effluent. During some tests samples were also taken at the aeration tank effluent. DO measurements were made periodically in the aeration tank.

To establish the normal loading variations a 24 hour sampling analysis was performed on the CM system on January 18, 1971. The results of this testing are illustrated in Figure 50.

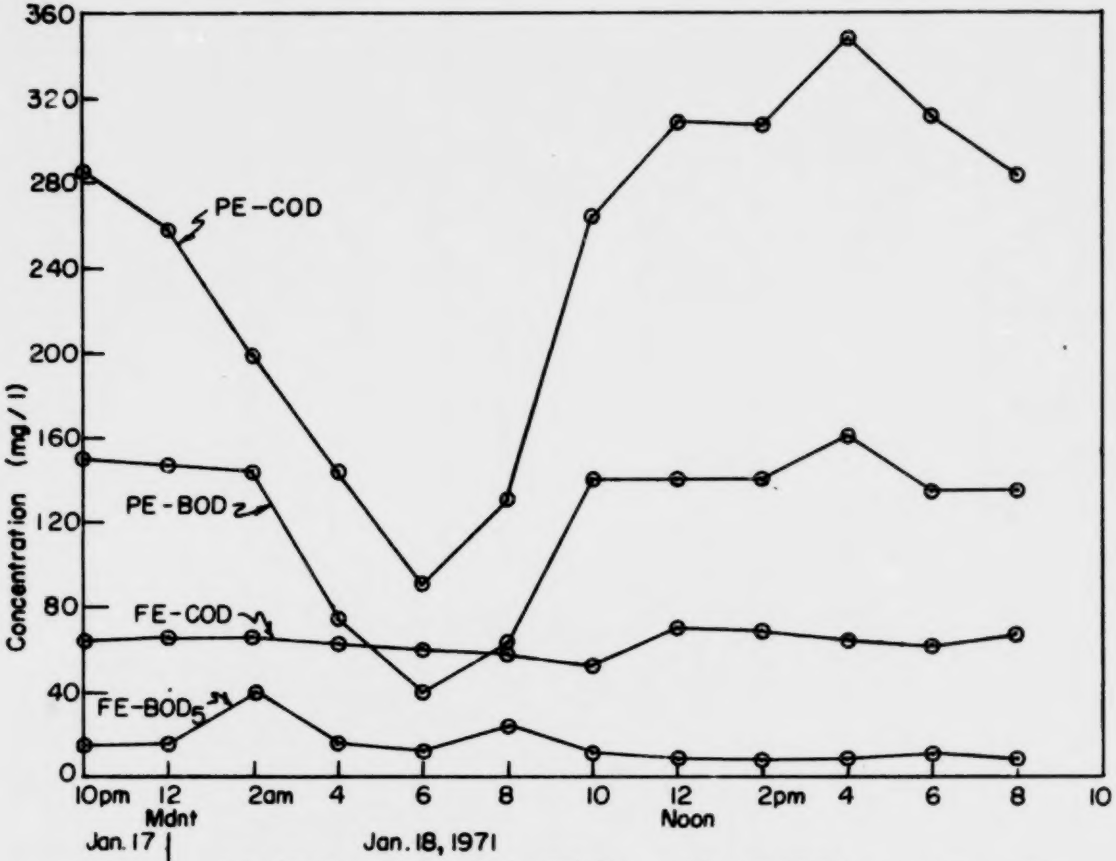
A source of high strength waste was located at the Kent Cheese Plant in Kent, Illinois. The cheese plant agreed to supply the Freeport Waste Treatment facility with 4500 gallon truck loads of whey from the cheese making process. The organic strength of the whey was estimated at 65,000 mg/l BOD_5.

During a shock load the flow entering the plant was bypassed around the grit chamber and primary settling tanks. The whey was dumped from a semi-truck into a manhole adjacent to the plant. Here it mixed with the incoming flow from the intercepter system and was then pumped directly to the aeration tanks.

Three shock load tests will be considered. During the first one (designated "Shock Load No. 2") the flow was split equally to each system. During the next test all the flow was directed into aeration tanks 2 and 4, the PF system (designated "No. 3"). Finally the last test consisted of applying all the flow to aeration tanks 1 and 3, the CM system (designated "No. 5").

Shock Load No. 2

Inlet flow for this test was as shown in Figure 1. The diffuser pattern for CM was as diagrammed in the " 198 total diffuser" diagram in Figure 5, and for PF as shown



24 HOUR PLANT EVALUATION - CM SYSTEM - **Figure 50**

in the lower portion of Figure 6. The average total sewage flow rate was 5.2 mgd.

The whey was applied over about a one half hour period and resulted in a sharp increase in loading as reflected by BOD, COD, and SS values designated as primary effluent in Figures 51,52, and 53 (actually "primary effluent" is a misnomer since the primaries werebeing by-passed). Final effluent No. 1 refers to the CM system and FE No. 2 to the PF system in these figures.

As would be expected, the peak BOD and COD concentrations occurred sooner in the CM system than the PF:

Parameter	Time to Peak in Effluent (hrs) CM	PF
BOD	2.25	4.50
COD	3.50	4.50
TSS	6.50	9.50

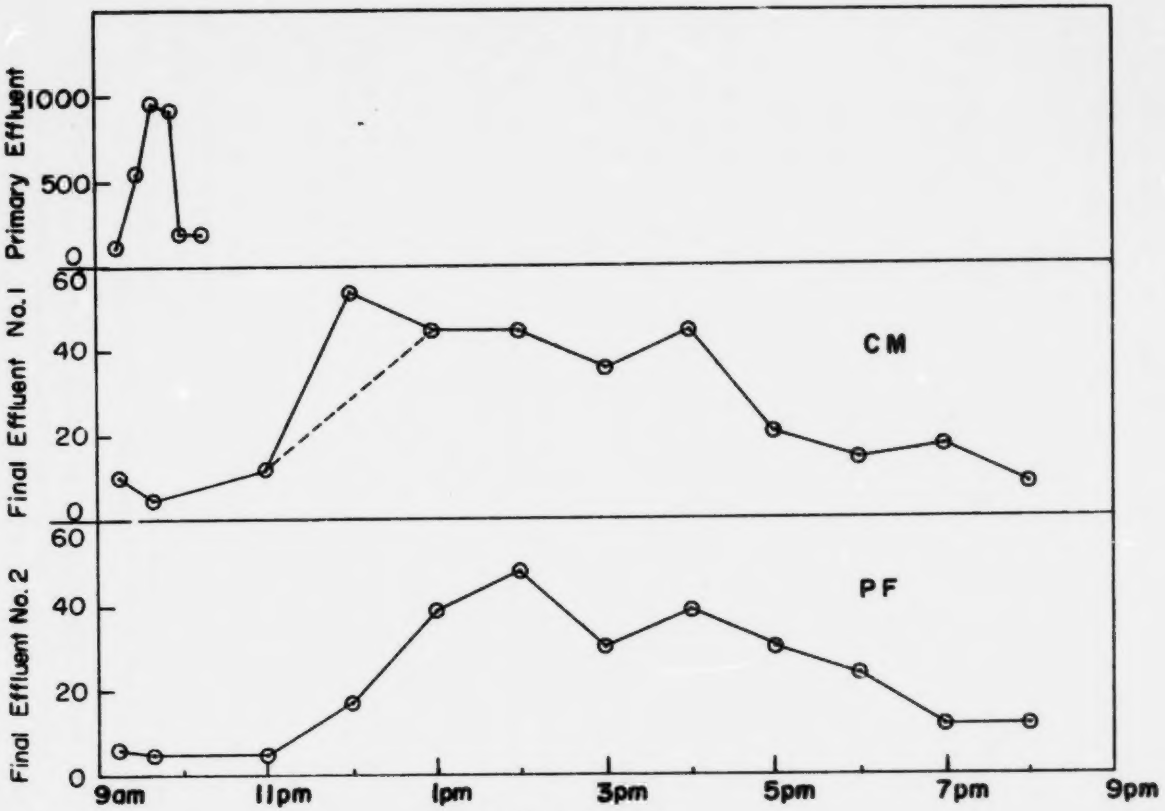
The tracer tests showed that the peak organic concentration would occur almost immediately at the AT effluent of the CM tanks, and the peak for PF should occur at the nominal T. Using the COD values, the detention time in the FSTs would be 3.5 hours. Nominal detention time would be 4.9 hours. On the basis of BOD data, T in the FSTs would be 2.25 hrs.

Assuming this to be true, and that T in the PF ATs is 2.4 hours, this would predict time-to-peak for the final effluent to be 4.65 hours. This compares favorably with the 4.5 hours shown above for BOD.

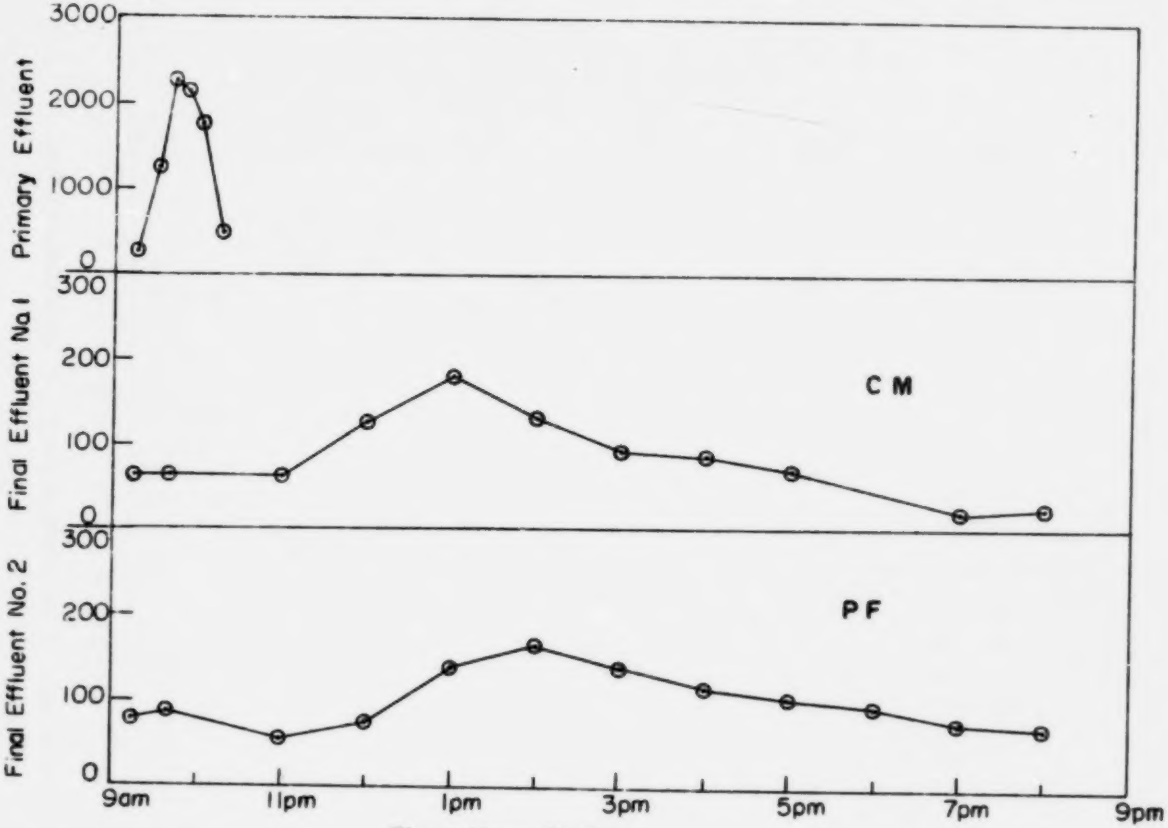
The point is that time-dependent organic tests bear out the tracer study results. Secondly, the data indicates that the average detention time of the final effluent portion of the influent flow to the FSTs is something less than the nominal detention time.

Peak values of final effluent BOD, COD and SS were very similar for each system. Actual mass removed for each parameter for this test was not calculated. Due to the similarity of response, it was decided to do another shock load at increased loading and decreased detention time.

During the test DOs were also measured in each of the aeration tanks. For the PF system, measurements were taken at the .25 and .75 points along the length. The results for PF are shown in Figure 54 and for CM in Figure 55. Of sig-

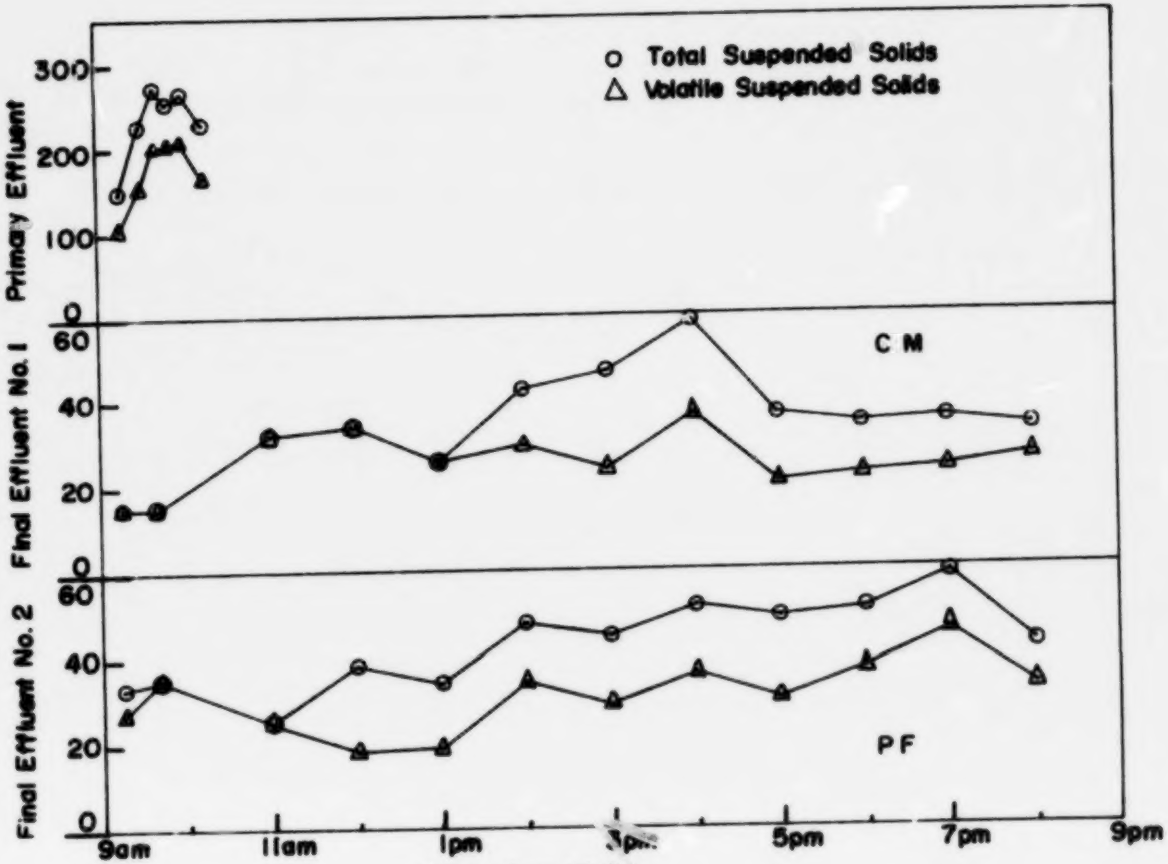


BOD$_5$ DATA - SHOCK LOAD NO. 2
Figure 51

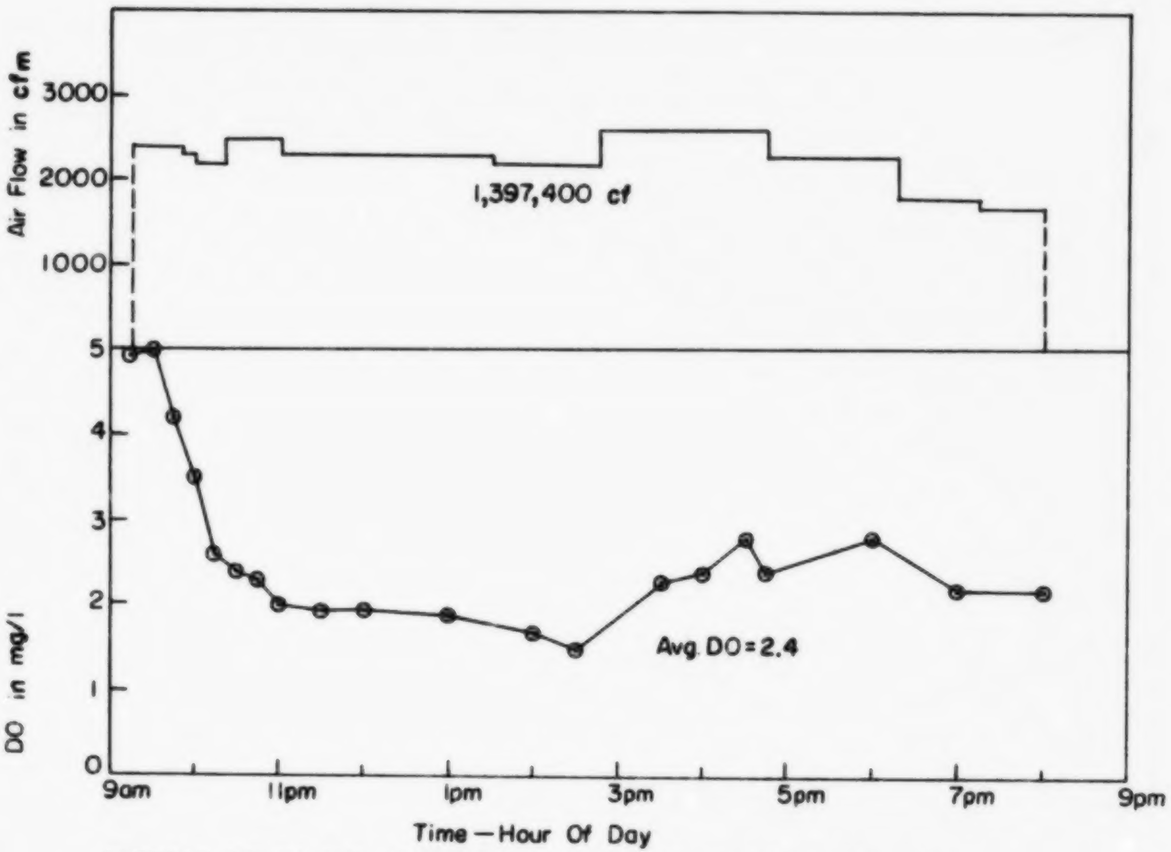


COD DATA — SHOCK LOAD NO.2

Figure 52

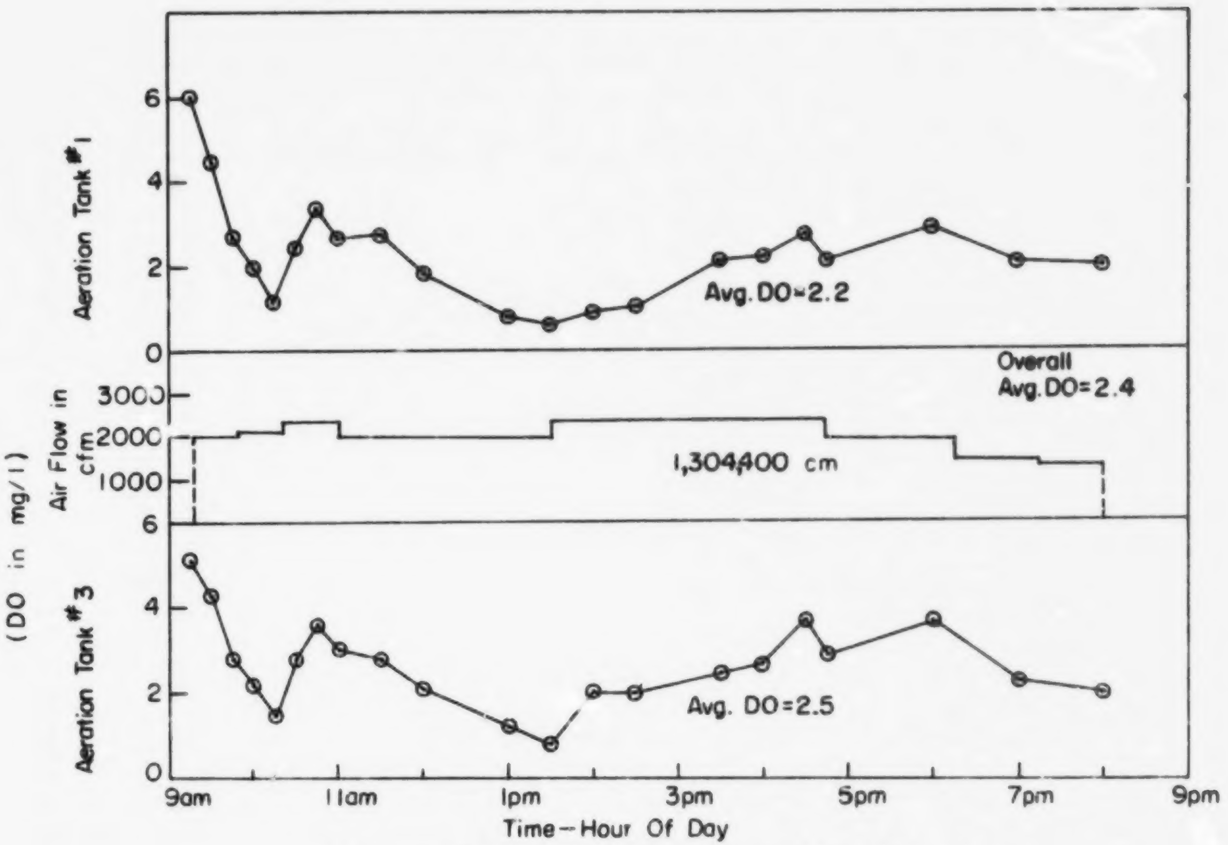


SOLIDS DATA—SHOCK LOAD NO.2
Figure 53



AVERAGE DO IN **PF SYSTEM** versus TIME — SHOCK LOAD NO. 2

Figure 54



DO versus TIME—COMPLETE-MIX—SHOCK LOAD NO. 2

Figure 55

nificance here is the fact that for the same organic load, the same sewage flow rate, and the same DO, more air was required for the PF system than the CM system (1,397,400 ft^3 as compared with 1,304,400 ft^3).

Shock Load No. 3

During this test the entire flow was directed into the PF system using the same inlet points and diffuser configuration described for Shock Load No. 2. The sewage flow rate averaged 4.0 mgd resulting in a nominal T of 1.58 hours.

To increase the organic load two whey dumps were used. The load peaks in terms of BOD, COD and SS are shown in the aeration tank influent graphs in Figures 56,57 and 58.

To separate the effects of aeration tank and final settling tank operation, samples were taken at the effluent of the PF aeration tank system. Each sample was filtered and then tested for BOD and COD. These results are shown in the center graph in Figures 56 and 57.

To carefully evaluate treatment efficiency, the mass of BOD and COD passing each sampling point was calculated using the sample data and the average sewage flow rate. The masses were determined for the period 9:30 am to 4 pm or 6.5 hours. Table 14 contains a summary of these calculations.

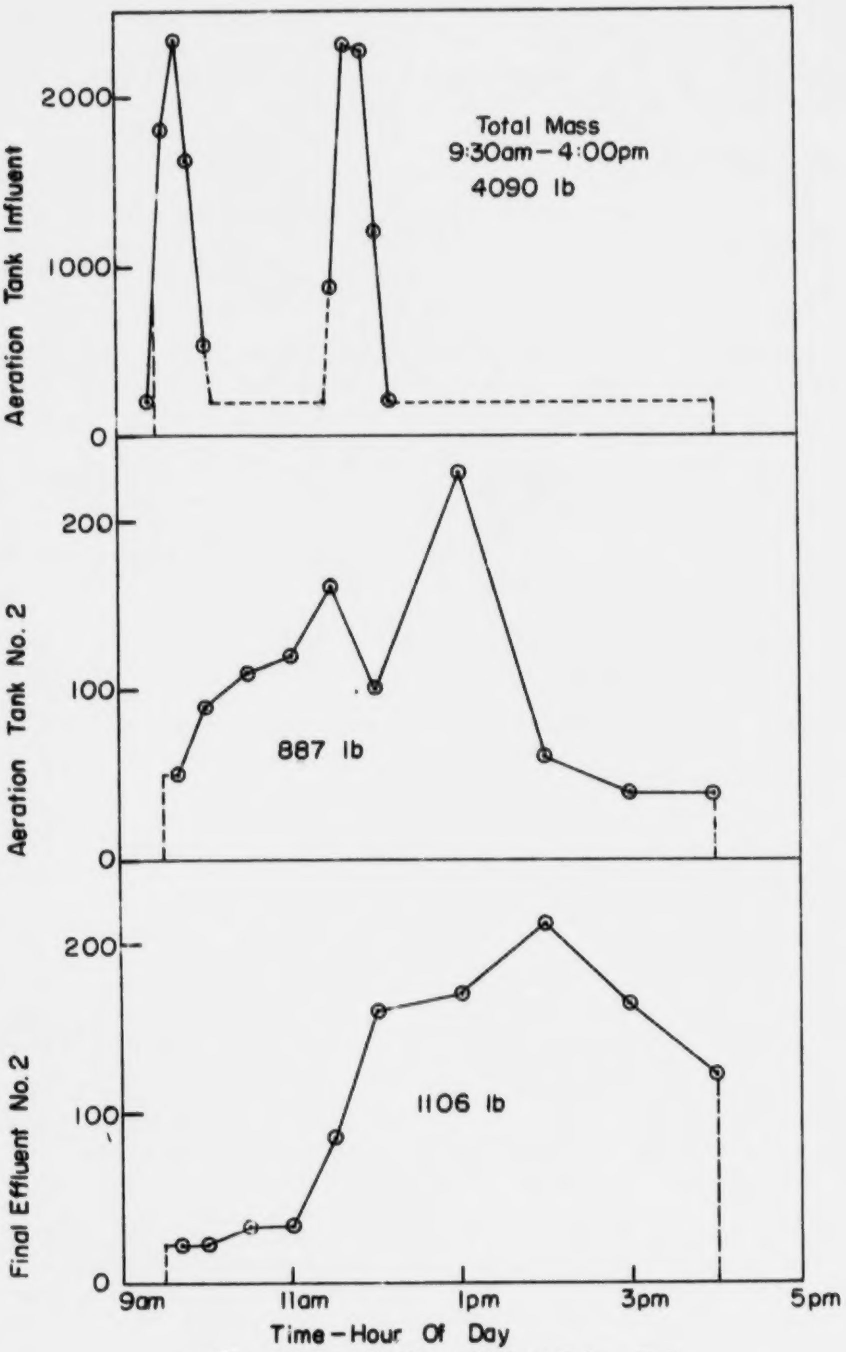
Air flow and DO measurements were taken. Again, the .25 and .75 points were chosen for DO measurement. Figure 59 shows the results of this testing.

Shock Load No. 5

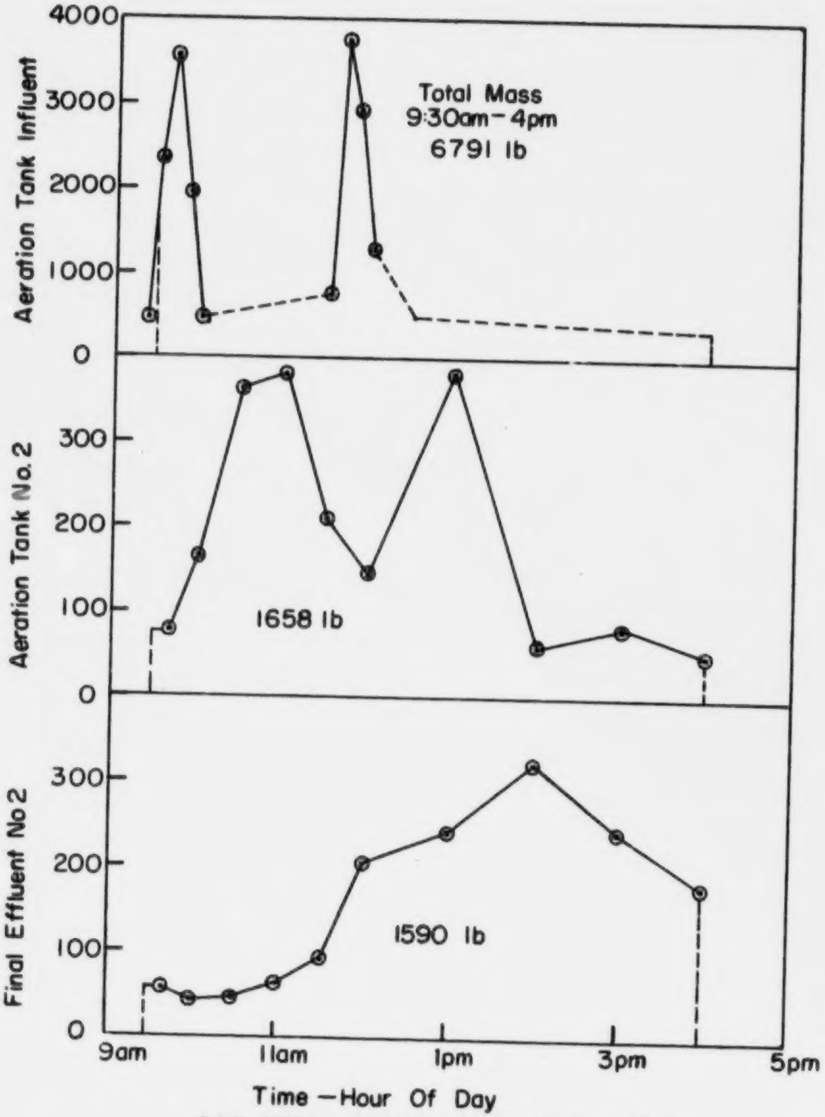
During this test, the entire flow was directed into the CM system using the same sewage and return sludge inlet points described for Shock Load No. 2. The diffuser configuration was changed, based on the results of the final tracer test, to the system using the over balance of diffusers towards the effluent side of the tank as diagrammed in Figure 5.

Problems with hauling the whey prevented applying a double shock load as was done in No. 3, so only a single aeration tank influent peak of BOD, COD and SS is shown in the upper graph of Figures 60,61 and 62.

The average sewage flow rate was 3.9 mgd as depicted in the upper graph in Figure 63.

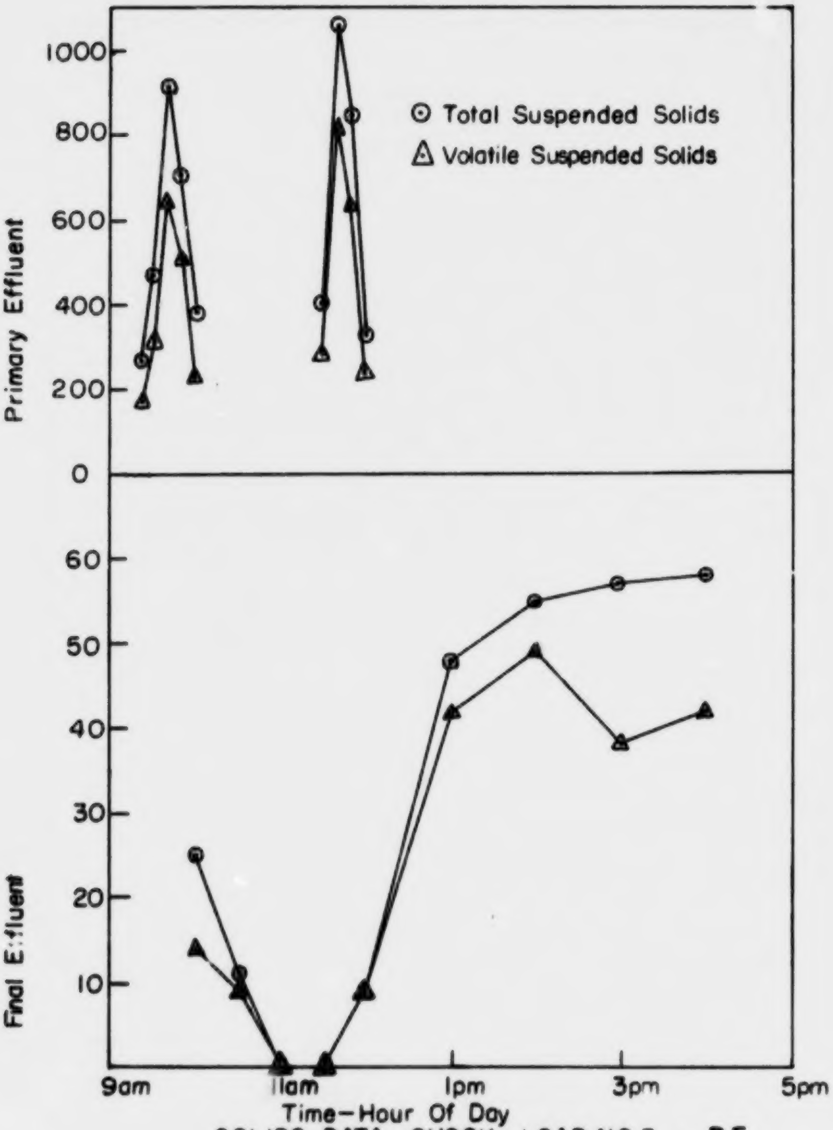


BOD DATA – SHOCK LOAD NO. 3 – P F
Figure 56

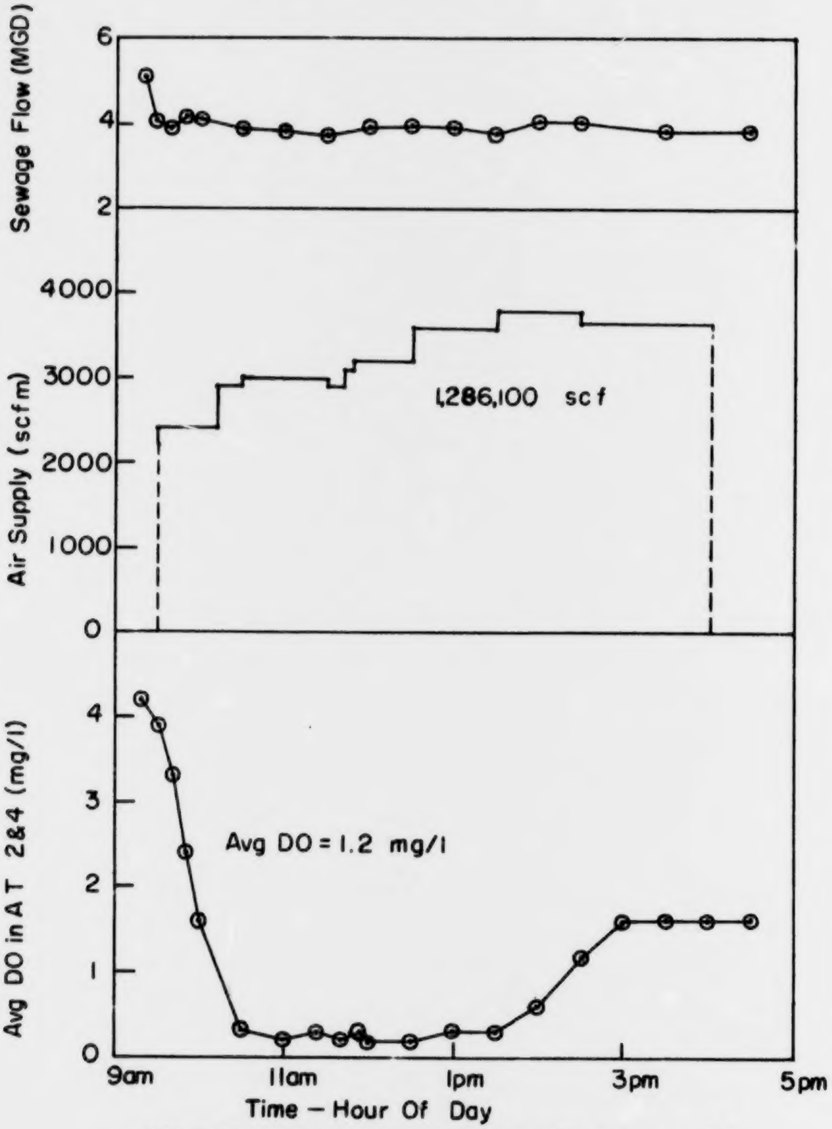


COD DATA— SHOCK LOAD NO. 3 - P F

Figure 57

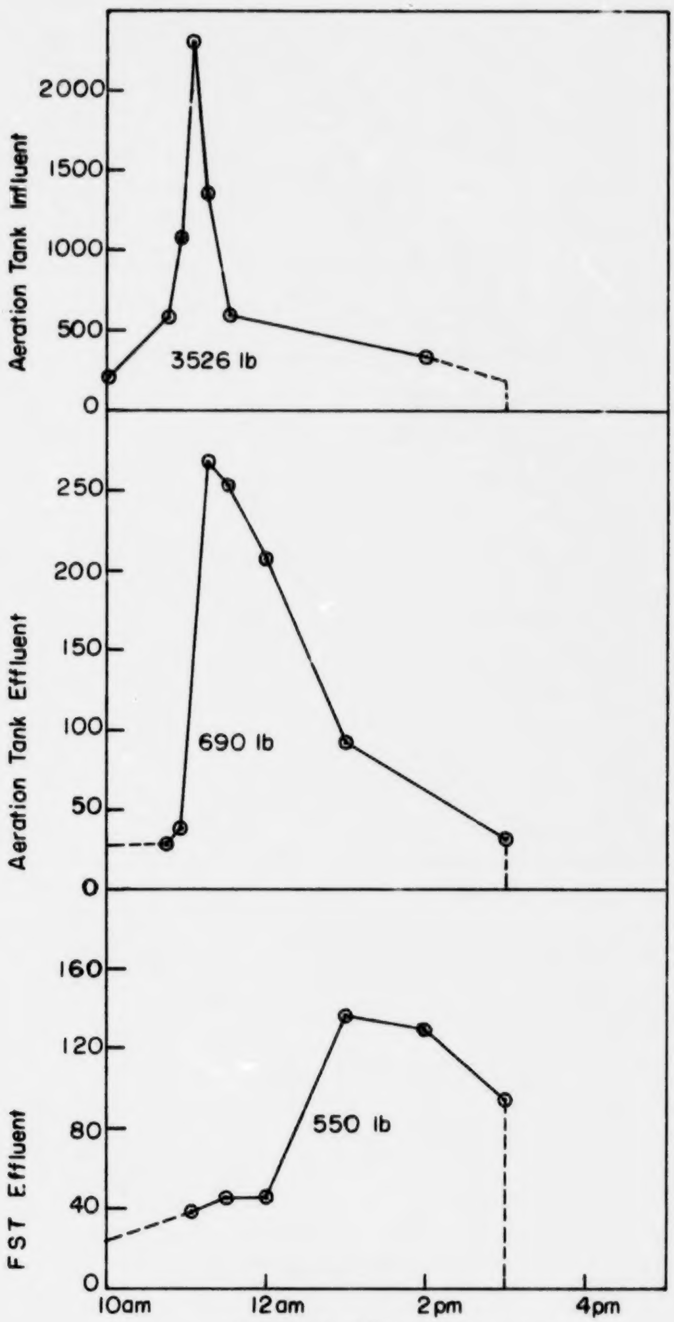


SOLIDS DATA – SHOCK LOAD NO. 3 – PF
Figure 58

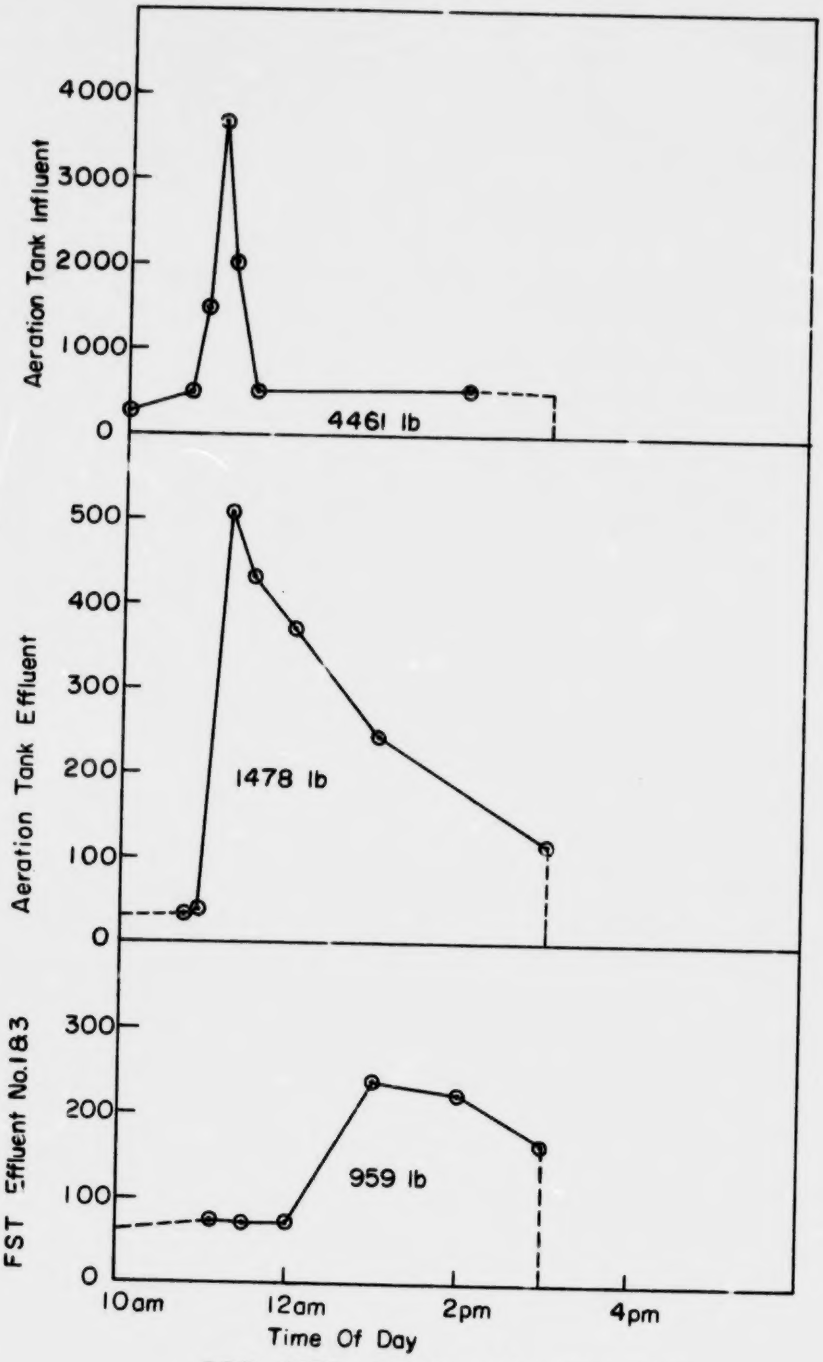


AIR & SEWAGE FLOW-SHOCK LOAD NO. 3 - PF

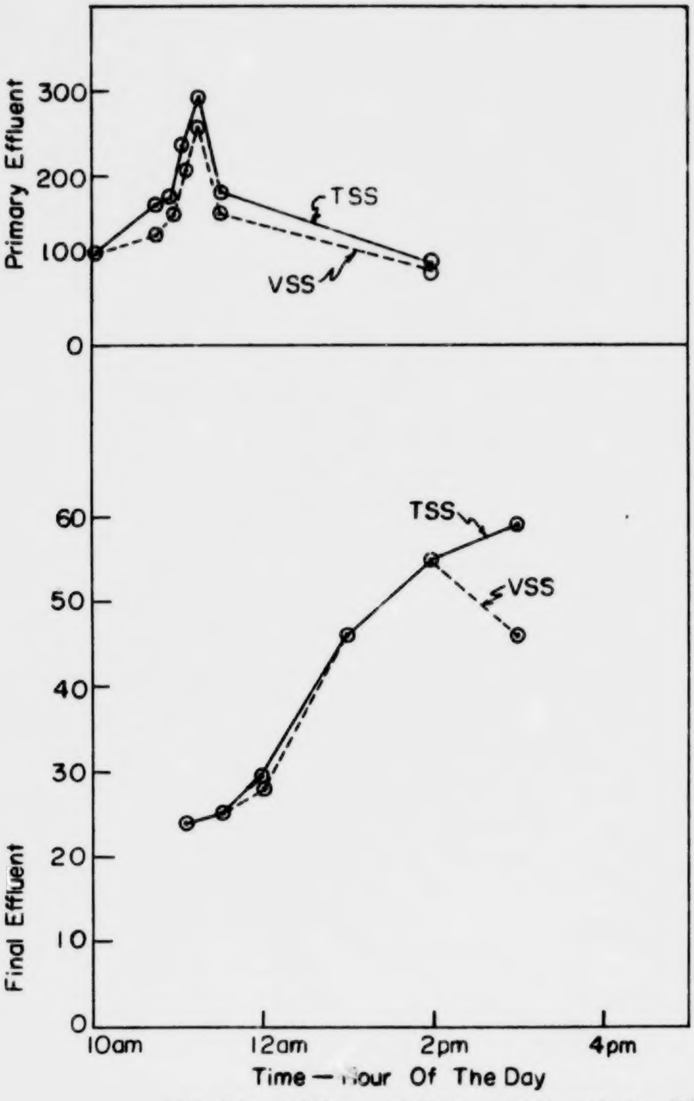
Figure 59



BOD$_5$ DATA- SHOCK LOAD NO. 5 — Figure **60**
CM



COD DATA-SHOCK LOAD NO.5-C M
Figure 61



SOLIDS DATA — SHOCK LOAD NO. 5 - CM

Figure 62

Treatment efficiency was again evaluated by calculating masses of BOD and COD entering and leaving the various points in the system. These calculations are summarized in Table 14.

Air flow and DO were measured periodically. As in all the tests, the air flow was adjusted to attempt to maintain a DO of 2 mg/l in each tank. A sharp drop in DO was noted (refer to Figure 63) during the time the shock load was applied. Such a drop was also experienced in the preceding tests. A discussion of system DO response follows in the next section.

Comparison of Shock Loads No. 3 and No. 5

The comparison data is summarized in Table 14. Each load was applied at about the same flow rate (4 mgd) yielding a nominal T of 1.6 hours. F/M loading was essentially the same (2.4 to 2.5).

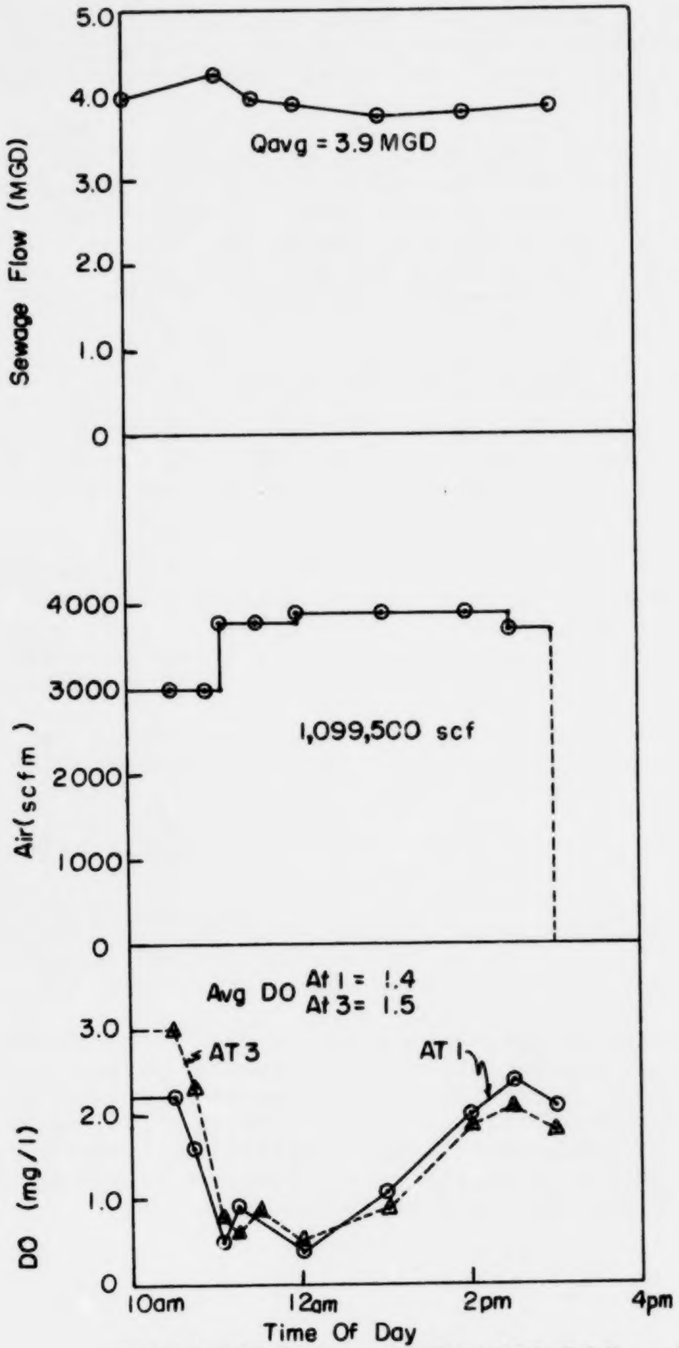
These tests indicate that the CM system yields more effective air utilization. First, comparing Figures 59 & 63, the drop in DO was not as severe nor of such long duration in the CM system. Second, Table 14 shows that the scf air per pound of BOD_5 removed was less for the CM tanks.

Removal percentages were somewhat better for CM. On the basis of aeration tank BOD_5 removal, CM was 80.4% compared with 78.3% for PF.

However COD removal data showed an opposite balance of 66.9% for CM and 75.6 for PF.

The CM did show better overall removal for both COD and BOD_5. It is interesting to note that CM effected considerable removal in the FSTs where the PF did not.

On the basis of the above data it can be concluded that CM does provide for improved treatment during periods of extreme shock loads involving a decreased detention time and an increased organic load. However, it should also be stated that under more normal detention times, e.g., "Shock Load No. 2", no significant difference in system performance between CM and PF could be determined.



AIR & SEWAGE FLOW — SHOCK LOAD NO.5 — Figure 63

COMPARISON SHOCK LOADS NO. 3 & NO. 5
PLUG-FLOW VERSUS COMPLETELY MIXED
TABLE 14

Parameter		(#3) Plug-Flow	Completely-Mixed (No. 5)	Comments
Test Period		6.5 hrs	5 hrs	Total air: PF = 1,286,100 scf CM = 1,099,500 scf Total lb BOD_5: PF = 4090 lb CM = 3526 lb DO was depressed to 0.2 for 2 hrs in PF. It only dropped to a low of .4 for about .5 hr in CM. The avg DO in PF would be lower except that initial values prior to the shock load were quite high 7-8 mg/l. Note: removal effected FSTs for BOD & COD in CM system.
Avg. Qs		4.0 mgd	3.9 mgd	
Detention Time		1.58 hrs	1.62 hrs	
$\frac{\text{SCF}}{\text{lb BOD}_5\text{ remvd}}$		431	369	
Avg. D.O.		1.2	1.5	
Avg. F/M $\frac{\text{lb BOD}_5\text{/day}}{\text{lb MLVSS}}$		2.38	2.55	
Removal Percentages				
BOD_5	PE-AT	78.3	80.4	
	PE-FE	73.0	84.4	
	AT-FE	-	20.3	
COD	PE-AT	75.6	66.9	
	PE-FE	76.6	78.5	
	AT-FE	4.1	35.1	

SECTION IX

PILOT PLANT OPERATION

Objective and Scope

The main objective of this investigation was to run a pilot plant utilizing a complete-mix activated sludge system in parallel with an existing full scale plant to determine the value of a model plant as a tool in the design and operation of future treatment plant. The goals are enumerated as follows:

1. To compare results of a scale model activated sludge plant operating under homogeneous mixing conditions (complete-mix) with a full scale plant operating under similar conditions.
2. To attempt to vary the operating parameters of the pilot plant to obtain optimum treatment efficiency and then compare these established parameter values with parallel ones in the prototype.

It is intended to obtain design constants from the operation of both plants, compare results from one plant with the other, and then see how well the pilot plant constants would have accurately predicted full scale operation, if such a pilot plant were to be used in the design of a similar full scale system. To facilitate the determination of the various design constants and parameters involved, the following tests were performed in addition to the routine laboratory analysis:

1. Oxygen uptake tests on samples of mixed liquor from the pilot plant aeration tank
2. Oxygen transfer efficiency tests using tap water and final effluent
3. A dye-tracer test to determine the mixing conditions in the aeration tank.
4. A 24-hour grab sample evaluation

Since similar tests were also run on the full-scale plant, comparisons of results can be made.

Equipment, Operation and Testing

Equipment

The pilot plant was installed at the end of the primary settling tanks, just prior to the point where the primary effluent enters the tunnel to the aeration tanks. This insured, basically, the same waste to both plants.

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The pilot plant (see schematic diagram in Figure 64) consists of the following two main units:

1. A 6" x 10" x 8' -0" deep aeration tank column with a six inch saran wrapped diffuser at the bottom
2. A 13.5" diameter by 7'-11" average depth final settling tank equipped with a small rotary sludge scraper at the bottom and with glass panels for viewing sludge level.

Primary effluent was pumped to the inlet of the aeration tank, as was the return sludge from the final settling tank. The pumps used were of the following specifications:

Randolph positive displacement type Model 500 with

zero-max speed control
catalogue No. 2130
minimum capacity: 130 cc/hr rated against water
maximum capacity: 45 g/hr
tube material: tygon or rubber
tube size: 7/16" OD x 1/4" ID

Air was supplied from the laboratory compressor. A manometer (U-tube type), a thermometer (°F) and a rotameter (to measure air flow rate) were introduced in the air line. The rotameter was of the following specifications:

Fisher & Porter, Lab Crest Mark III Tri Flat

flowmeter kit
range: 0.4 cc/min to 23,400 cc/min

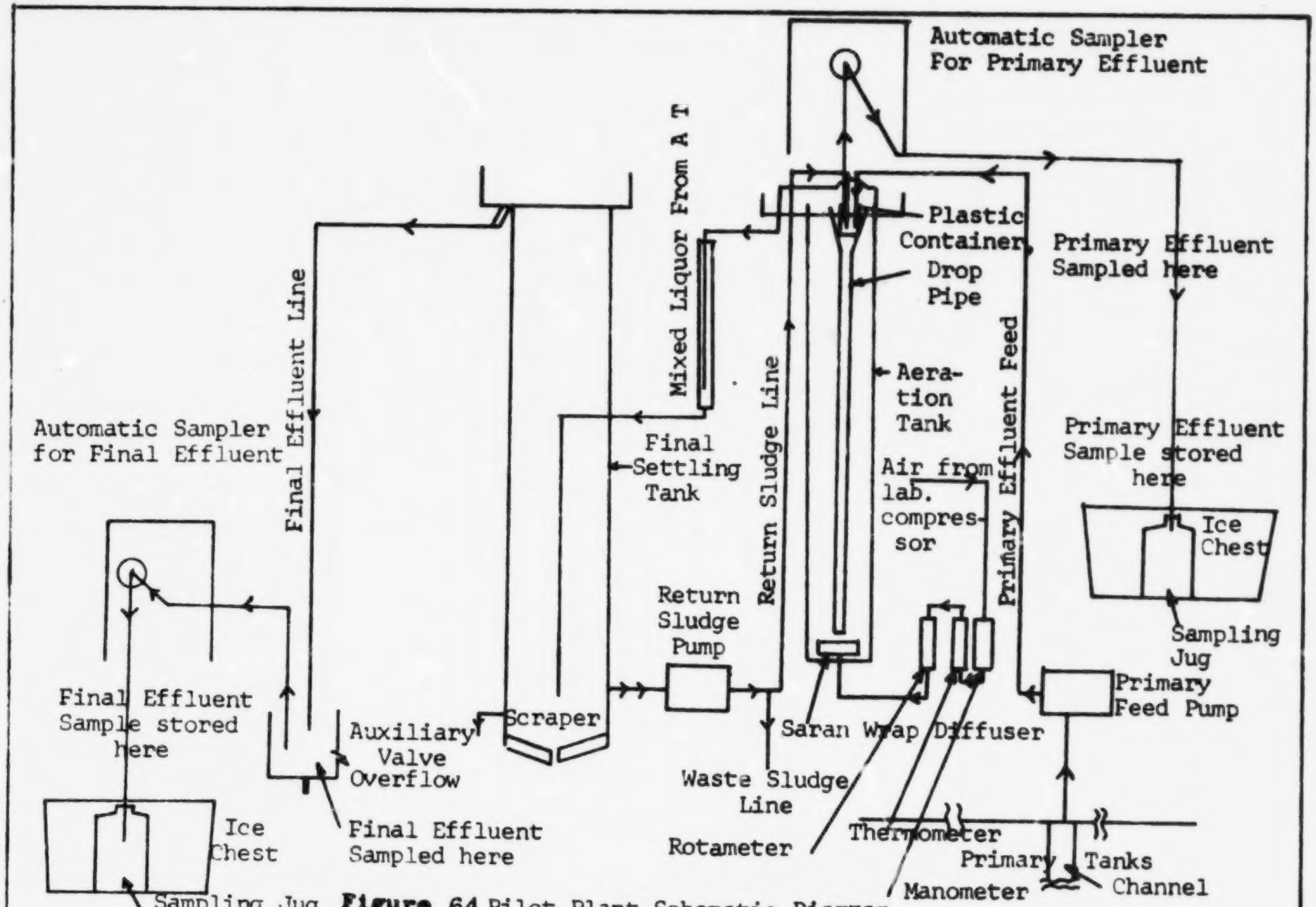
Two of the flowmeters were calibrated to give the air flow rates (cc/min) at prevailing temperature and pressure. The pilot plant equipment included two automatic sampling units manufactured by Brailsford and Company, New York and having the following specifications:

Model EV2 and EV3

Other items included sampling jugs and ice buckets to preserve the samples.

Operation

Primary effluent was pumped into the top of the aeration tank and into a drop pipe which extended down to a few inches above the diffuser. The mixed liquor flowed over a weir through a 1.5" pipe to the settling tank, where it was discharged at the bottom. The settled solids were collected by a scraper, and the return sludge was pumped from the tank



Automatic Sampler
For Primary Effluent
Plastic
Container
Primary Effluent
Sampled here
Drop
Pipe
Mixed Liquor From A T
Final Effluent Line
Aeration
Tank
Primary Effluent Feed
Primary Effluent
Sample stored
here
Automatic Sampler
for Final Effluent
Final
Settling
Tank
Return Sludge Line
Air from
lab.
compressor
Ice
Chest
Return
Sludge
Pump
Sampling
Jug
Primary
Feed Pump
Final Effluent
Sample stored
here
Auxiliary
Valve
Overflow
Scraper
Saran Wrap Diffuser
Waste Sludge
Line
Ice
Chest
Final Effluent
Sampled here
Rotameter
Thermometer
Primary
Tanks
Channel
Manometer
Sampling Jug

Figure 64 Pilot Plant Schematic Diagram

bottom up to the aeration tank inlet.

Sludge wasting was done by opening a valve located just beyond the effluent point of the return sludge pump

The effluent from the final settling tank flowed over a circular V notched weir into a 3/4" rubber hose.

The primary feed rate was determined by measuring the effluent flow rate. A stopwatch and a graduated container were used to measure this flow rate as well as the return sludge rate which was measured just before entry into the aeration tank.

One important aspect of the whole pilot plant operation was sampling. It was decided that composite sampling would be most representative, especially since that was the method adopted in the full scale operations. During the initial part of the pilot plant operation, sampling was done manually. Every hour 100 ml of primary effluent was picked up from the trough in which the influent line of the primary feed pump was submerged. An equal amount of final effluent was collected as it came out of the effluent rubber hose. The composite samples were collected in jugs which were refrigerated. However, it was felt that this system was prone to human errors, and it was decided to install two automatic samplers. One was mounted at the point where the primary flow discharged into the aeration tank and the other at the outlet hose of the final settling tank. Plastic containers were placed at these two points, the liquid level being maintained by sizing a hole in the bottom of each to allow flow to go through with a sufficient head. In this manner the intake of the automatic sampler was always under the liquid surface. The samplers were set to remove about a gallon of sample every 24-hour period. They discharged into plastic jugs which were kept in insulated boxes in which ice was placed each morning. In this way, the samples were maintained at approximately 40° F and the entire sampling system was automatic.

The pilot plant was essentially operated in two phases:

1. In parallel with the complete mix side of the full scale
2. At various periods of constant detention time

There were four periods of parallel operation. During each of these periods an attempt was made to control hydraulic flows (million gallons/day versus l/hr) in both plants to yield similar average detention times. The mixed liquor concentration was also kept the same so that each plant would be similarly loaded. It was originally intended to maintain the same return sludge to feed ratio. However, due to resulting operational problems in the pilot plant, the ratio had to be increased to 0.5 from the original 0.3, which was maintained in the full scale operations.

Flow equalization was fairly well achieved by matching the flows exactly in each plant every two hours, starting at 8 am on each day. Return sludge rates were correspondingly changed. At 4 pm, or in some cases as late as 10:30 pm, the flow through the pilot plant was adjusted to a rate which would correspond to the average flow through the full scale plant between then and 8 am the next morning. Figure 65 gives an idea of flow variations in both plants. It is a typical representation.

Five ten-day periods were chosen for the constant detention operation. In each period the detention time was held constant. The detention times chosen were from a nominal of 5 hours down to 1 hour. It was intended originally to achieve three levels of MLSS in each period (1,000, 2,000 and 3,000 mg/l). However, owing to limitations of time in each period, only one or two distinct solids levels were obtained.

In each operational phase, it was intended to maintain a return sludge to feed ratio of 0.5 and a DO level of 2.0 mg/l.

Testing

Routine laboratory tests were run on pilot plant composite samples in the same manner as on the full scale composite samples. The following tests were run daily:

1. Total suspended solids (TSS)
2. Volatile suspended solids (VSS)
3. 5-day BOD on unfiltered samples (BOD5UF)
4. 5-day BOD on filtered samples (BOD5F)
5. Chemical oxygen demand on unfiltered samples (CODUF)
6. Chemical oxygen demand on filtered samples (CODF)

Mixed liquor and return sludge samples were withdrawn at the same time every day. Besides the solids test on both samples, the half-hour settling test was also conducted on the mixed liquor samples.

Other measurements included:

1. Aeration tank DO using a YSI oxygen probe
2. Mixed liquor temperature
3. Air measurements that included flow rate and temperature and pressure readings

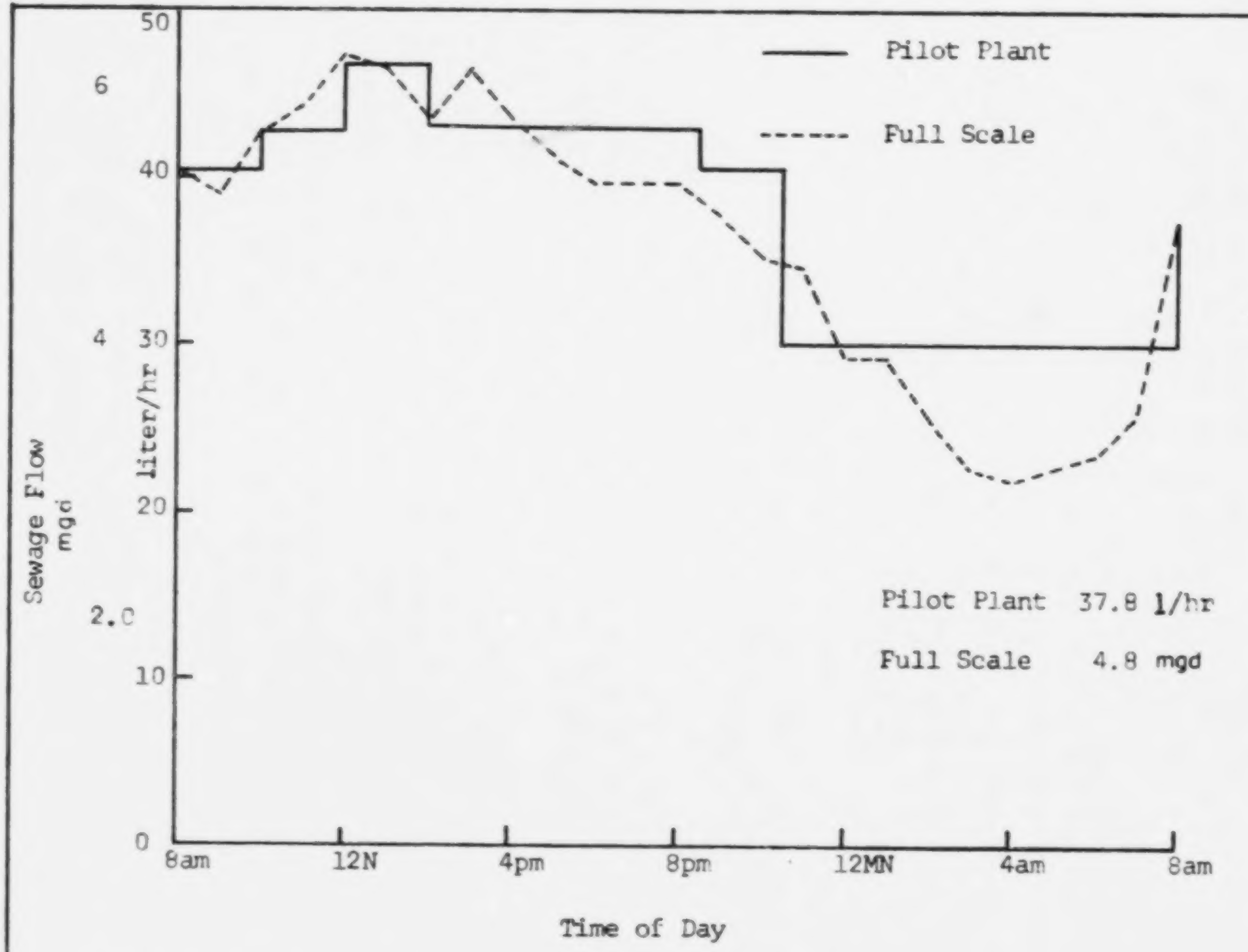


Figure 65 Typical 24 Hour Flow Pattern During Parallel Evaluation

Feed and return sludge flow rates were checked periodically, especially during the parallel operation when it was necessary to match flows with the full scale plant. During the constant T operation, flows were checked daily and adjusted as required.

Some special tests were conducted, the results of which were incorporated with the operating data. They were:

1. Oxygen uptake tests
2. Dye tracer test
3. Oxygen transfer efficiency test

All the laboratory tests were done in accordance with Standard Methods[14] with the following alternate methods being adopted. In the solids test an asbestos fiber mat was used to filter composite primary effluent and final effluent samples, while mixed liquor samples were filtered through 1 inch diameter Gelman Type A glass fiber filters. All DO measurements were taken with a YSI DO probe.

Pilot Plant Operational Problems

The only major problem involving malfunctioning of equipment was caused by the sludge scraper mechanism. The original motor failed during the initial stages of operation. This resulted in a very low concentration of return sludge, since the scraper at the bottom of the settling tank worked intermittently. Replacement motors were not powerful enough to work the scraper. Eventually, the whole mechanism was replaced by a new one which was different from the original in that the new drive turned at 1 rpm instead of 1/3 rpm. This caused the solids at the bottom of the final settling tank to be in continuous agitation. Though return sludge concentrations were quite normal, it is suspected that the faster drive did cause a certain amount of solids carryover resulting in higher final effluent suspended solids values.

The primary feed pump and the return sludge pump worked with reduced efficiency at low flow rates. Also, the hoses in the flow streams tended to clog because of solids deposition. Even though the hoses were replaced periodically, solid deposition was unavoidable. Thus there was a definite loss of solids in this manner.

From the operational point of view, it was felt that the sampling program was not ideally representative. Because of its size, the system was very responsive to slight changes, and a continuously steady-state operation seemed unlikely. It is felt that a testing program repeated at least 3 times daily (as in the full scale operation) would present more meaningful data, expecially during the constant detention time (T) operation. It was found to be extremely difficult to vary the MLSS concentrations in a reasonable period of time when operating at long detention times. It was because of this reason that a study of optimum MLSS and detention time combination was not entirely successful.

Certain improvements in results during the operation were noticed and the following observation can be made:

1. Automatic sampling was much more representative and efficient than manual sampling, as far as composite sampling is concerned. Many errors were avoided by automation.
2. There was definitely better operation with rubber hoses than with tygon tubing, especially for lines leading into and out of the feed and return sludge pumps. Rubber hosing decreased flow pulsations considerably and had a longer life.

Results

In this chapter the results are presented first. However, for each period of operation, removal efficiencies, loading and other related parameters have been summarized. The results of the various special tests conducted are also reported. Interpretation of the results and operational data are presented in the second half of this chapter.

Introduction

The results from the parallel operation are given for each period separately in Tables 15 to 18. Corresponding values for each parameter for the full scale operation are also given in these tables.

Tables 19 to 23 summarize each period of constant detention time individually, giving corresponding full scale values as before. Table 24 is a summary of the pilot plant constant T operation, and Table 25 is a similar table for the full scale operation.

For comparing pilot-full scale operation, consider each period in turn, taking parallel operation first.

TABLE 15

Pilot-Full Scale Comparison

June 15-30, 1971

Parameter	Pilot Plant	Full Scale
BOD Removal % PEUF to FEF	84	81
BOD Removal % PEUF to FEUF	60	58
COD Removal % PEUF to FEF	70	78
COD Removal % PEUF to FEUF	58	63
Solids Removal %		
TSS	68	69
VSS	70	68
$\frac{\text{Lb BOD}}{\text{day lb}}$ MLVSS	0.65	0.72
$\frac{\text{SCF air}}{\text{Lb BOD}}$	1345	908
$\frac{\text{Lb BOD}}{1000\ \text{ft}^3}$	55	68
Detention Time, hrs	2.2	2.1
Loading on FST		
Hydraulic, gpd/ft^2	348	800
Solids, lb/day/ft^2	7.0	17.3
Average: Flow **(.000267** mgd **)**	42.1 lit/hr.	3.1 **mgd**
MLSS	2427 mg/l	2622 mg/l
MLVSS	1332 mg/l	1524 mg/l
PEUF BOD$_5$	83.0 mg/l	93.0 mg/l

TABLE 16

Pilot-Full Scale Comparison

July 1-11 & 26-31, 1971

Parameter	Pilot Plant	Full Scale
BOD Removal % PEUF to FEF	87.1	89.4
BOD Removal % PEUF to FEUF	69.5	72.7
COD Removal % PEUF to FEF	72.9	75.5
COD Removal % PEUF to FEUF	65.3	61.9
Solids Removal %		
TSS	67.7	75.6
VSS	67.9	75.6
$\frac{\text{Lb BOD}}{\text{day lb}}$ MLVSS	0.63	0.82
$\frac{\text{SCF air}}{\text{Lb BOD}}$	678	842
$\frac{\text{Lb BOD}}{\text{1000 ft}^3}$	51	56
Detention Time, hrs	2.9	3.3
Loading on FST		
Hydraulic, gpd/ft ^{2}	286	494
Solids, lb/day/ft ^{2}	4.9	6.6
Average: Flow **(.000211 mgd)**	33.3 lit/hr	1.94 mgd
MLSS	2066 mg/l	1590 mg/l
MLVSS	1296 mg/l	1100 mg/l
PEUF BOD_5	98 mg/l	123 mg/l

TABLE 17

Pilot-Full Scale Comparison

July 12-25, 1971

Parameter	Pilot Plant	Full Scale
BOD Removal % PEUF to FEF	88.9	92.9
BOD Removal % PEUF to FEUF	73.4	71.2
COD Removal % PEUF to FEF	76.5	82.0
COD Removal % PEUF to FEUF	67.8	76.4
Solids Removal %		
TSS	70.5	86.2
VSS	68.7	86.8
$\frac{\text{Lb BOD}}{\text{day lb MLVSS}}$	0.60	0.57
$\frac{\text{SCF air}}{\text{Lb BOD}}$	686	963
$\frac{\text{Lb BOD}}{\text{1000 ft}^3}$	38	45
Detention Time, hrs	3.0	3.2
Loading on FST		
Hydraulic, gpd/ft ^{2}	268	509
Solids, lb/day/ft ^{2}	3.3	8.2
Average: Flow (.000202 mgd)	31.8 lit/hr	2.0 mgd
MLSS	1459 mg/l	1926 mg/l
MLVSS	982 mg/l	1272 mg/l
PEUF BOD_5	76.0 mg/l	96 mg/l

TABLE 18

Pilot-Full Scale Comparison

August 10-19, 1971

Parameter	Pilot Plant	Full Scale
BOD Removal % PEUF to FEF	91	89.4
BOD Removal % PEUF to FEUF	79.1	79.1
COD Removal % PEUF to FEF	75.9	82.9
COD Removal PEUF to FEUF	63.8	72.2
Solids Removal %		
TSS	69.7	73.9
VSS	69.7	71.2
$\frac{\text{Lb BOD}}{\text{day lb}}$ MLVSS	0.41	0.49
$\frac{\text{SCF air}}{\text{Lb BOD}}$	1750	911
$\frac{\text{Lb BOD}}{\text{1000 ft}^3}$	33	43
Detention Time, hrs	2.7	2.7
Loading on FST		
Hydraulic, gpd/ft^2	337	580
Solids, lb/day/ft^2	6	10.5
Average: Flow (.000223 mgd)	35.18 lit/hrs	2.32 mgd
MLSS	2088 mg/l	2147 mg/l
MLVSS	1273 mg/l	1398 mg/l
PEUF BOD_5	60 mg/l	79 mg/l

TABLE 19

Constant Detention Time Operation
Pilot-Full Scale Comparison

5-Hour Nominal Detention Time

Parameter	Pilot Plant	Full Scale January 1-31	Full Scale February 1-12
BOD Removal % PEUF to FEF	93.7	89.6	92.1
BOD Removal % PEUF to FEUF	74.3	85.9	87.4
COD Removal % PEUF to FEF	85.3	84.7	83.7
COD Removal % PEUF to FEUF	70.2	75.6	75.9
Solids Removal %			
TSS	79.8	82.0	77.4
VSS	79.6	82.5	72.8
$\frac{\text{Lb BOD}}{\text{day lb MLVSS}}$	0.29	0.26	0.19
$\frac{\text{SCF air}}{\text{Lb BOD}}$	2105	1070	1180
$\frac{\text{Lb BOD}}{1000 \text{ ft.}^3}$	24	41	37
Detention Time, hrs.	5.06	4.9	5.1
Loading on FST			
Hydraulic, gpd/ft ^{2}	180	331	316
Solids, lb/day/ft ^{2}	3.4	10.5	11.7
Average: Flow (.000119 mgd)	18.7 lit/hrs	1.30 mgd	1.24 mgd
MLSS	2271 mg/l	3790 mg/l	4450 mg/l
MLVSS	1343 mg/l	2500 mg/l	3080 mg/l
PEUF BOD_5	84 mg/l	135 mg/l	127 mg/l

TABLE 20

Constant Detention Time Operation
Pilot-Full Scale Comparison

4-Hour Nominal Detention Time

Parameter	Pilot Plant	Full Scale February 13-17	 March 12-17
BOD Removal % PEUF to FEF	91.9	90.4	92.2
BOD Removal % PEUF to FEUF	69.1	90.4	81.1
COD Removal % PEUF to FEF	85.4	84.5	78.8
COD Removal % PEUF to FEUF	81.1	90.8	74.0
Solids Removal %			
TSS	79.5	87.4	91.0
VSS	78.9	85.6	88.7
$\frac{\text{Lb BOD}}{\text{day lb MLVSS}}$	0.35	0.24	0.21
$\frac{\text{SCF air}}{\text{Lb BOD}}$	1327	990	1060
$\frac{\text{Lb BOD}}{1000\ \text{ft}^3}$	33	49	34
Detention Time, hrs	4.04	4.1	4.0
Loading on FST			
Hydraulic, gpd/ft^2	224	393	405
Solids, lb/day/ft^2	4.3	15.1	15.8
Average: Flow (000148 mgd)	23.4 lit/hrs	1.54 mgd	1.59 mgd
MLSS	2314 mg/l	4620 mg/l	4670 mg/l
MLVSS	1454 mg/l	3340 mg/l	2560 mg/l
PEUF BOD_5	92 mg/l	135 mg/l	90 mg/l

TABLE 21

Constant Detention Time Operation
Pilot-Full Scale Comparisons

3-Hour Nominal Detention Time

Parameter	Pilot Plant	Full Scale
BOD Removal % PEUF to FEF	91.5	89.7
BOD Removal % PEUF to FEUF	74.4	83.3
COD Removal % PEUF to FEF	79.5	79.1
COD Removal % PEUF to FEUF	72.5	69.5
Solids Removal %		
TSS	70.5	73.0
VSS	72.7	71.2
$\frac{\text{Lb BOD}}{\text{day lb}}$ MLVSS	0.4	0.33
$\frac{\text{SCF air}}{\text{Lb BOD}}$	960	790
$\frac{\text{Lb BOD}}{\text{1000 ft}^3}$	40	38
Detention Time, hrs	2.99	3.0
Loading on FST		
Hydraulic, gpd/ft ^{2}	304	530
Solids, lb/day/ft ^{2}	6.3	13.3
Average: Flow (.000200 mgd)	31.6 lit/hrs	2.08 mgd
MLSS	2485 mg/l	3000 mg/l
MLVSS	1607 mg/l	1850 mg/l
PEUF BOD_5	82 mg/l	78 mg/l

TABLE 22

Constant Detention Time Operation
Pilot-Full Scale Comparison

2-Hour Nominal Detention Time

Parameter		Pilot Plant	Full Scale
BOD Removal % PEUF to FEF		92.6	91.8
BOD Removal % PEUF to FEUF		65.3	87.7
COD Removal % PEUF to FEF		79.4	79.1
COD Removal % PEUF to FEUF		70.0	70.9
Solids Removal %			
TSS		70.7	71.4
VSS		66.1	70.2
$\frac{\text{Lb BOD}}{\text{day Lb}}$ MLVSS		0.69	0.51
$\frac{\text{SCF air}}{\text{Lb BOD}}$		564	430
$\frac{\text{Lb BOD}}{1000\ \text{ft}^3}$		70	82
Detention Time, hrs		1.98	2.2
Loading on FST			
Hydraulic, gpd/ft^2		451	724
Solids, lb/day/ft^2		8.3	22.5
Average:	Flow **(000302** mgd)	47.67 lit/hr.	2.84 mgd
	MLSS	2213 mg/l	3720 mg/l
	MLVSS	1610 mg/l	2560 mg/l
	PEUF BOD_5	95 mg/l	122 mg/l

TABLE 23

Constant Detention Time Operation
Pilot-Full Scale Comparison

1-Hour Nominal Detention Time

Parameter	Pilot Plant	Full Scale
BOD Removal % PEUF to FEF	89	91
BOD Removal % PEUF to FEUF	76	79
COD Removal % PEUF to FEF	N/A	81
COD Removal % PEUF to FEUF	N/A	74
Solids Removal %		
TSS	69	83
VSS	69	82
$\frac{\text{Lb BOD}}{\text{day lb}}$ MLVSS	0.74	1.00
$\frac{\text{SCF air}}{\text{Lb BOD}}$	505	310
$\frac{\text{Lb BOD}}{1000 \text{ ft}^3}$	167	210
Detention Time, hrs	1.1	1.1
Loading on FST		
Hydraulic, gpd/ft ^{2}	824	727
Solids, lb/day/ft ^{2}	38	24
Average: Flow **(000545** mgd**)**	86. lit/hr	2.85 mgd
MLSS	5568 mg/l	4850 mg/l
MLVSS	3627 mg/l	3310 mg/l
PEUF BOD_5	123 mg/l	156 mg/l

TABLE 24

Summary of Constant Detention Time Studies
Pilot Plant

Period	Raw Flow 1/hr	Det. Time	Primary Effluent						Final Effluent					
			BOD		COD				BOD		COD			
	Q	hrs T	F	UF	F	UF	TSS	VSS	F	UF	F	UF	TSS	VSS
August 20 to August 30	18.7	5.06	27	84	51	218	91	66	5	22	32	65	18	14
September 1 to September 10	23.4	4.04	38	92	47	204	94	74	7	28	30	39	19	16
September 11 to September 20	31.6	2.99	27	82	63	171	95	66	7	21	35	47	28	18
September 21 to September 30	47.7	1.98	28	95	52	170	75	59	7	33	35	51	22	20
April 27 to May 7	86.0	1.1	74	123	-	-	124	92	14	29	-	-	38	31

TABLE 24 (continued)

Period	Per Cent Removal (%) BOD PEUF-FEF	COD PEUF-FEF	TSS	VSS	F/M lb BOD / day, lb MLVSS	Mixed Liquor Temp ^{o}C
August 20 to August 30	93.7	85.3	79.8	79.6	0.29	≈ 23
September 1 to September 10	91.9	85.4	79.5	78.9	0.35	≈ 24
September 11 to September 20	91.5	79.5	70.5	72.7	0.40	≈ 21
September 21 to September 30	92.6	79.4	70.7	66.1	0.69	≈ 20
April 27 to May 7	89	-	69.0	66.0	0.74	≈ 16

TABLE 25

Summary of Constant Detention Time Studies
Full Scale Plant

Period	Raw Flow Q, mgd	Det. Time T, hrs	Primary Effluent						Final Effluent					
			BOD_5		COD				BOD_5		COD			
			F	UF	F	UF	TSS	VSS	F	UF	F	UF	TSS	VSS
January 1 to January 31	1.30	4.9	82	135	99	287	150	126	14	19	44	70	27	22
February 1 to February 12	1.24	5.1	77	127	98	245	115	92	10	16	40	59	26	25
February 13 to February 17	1.54	4.1	95	135	103	265	127	104	13	13	41	51	16	15
March 12 to March 17	1.59	4.0	39	90	66	208	199	124	7	17	44	54	18	14
March 26 to April 5	2.08	3.0	33	78	67	187	107	80	8	13	39	57	29	23
April 6 to April 22	2.84	2.2	81	122	72	206	126	94	10	15	43	60	36	28
April 24 to May 9	2.85	1.1	104	156	72	226	161	120	14	33	44	58	27	22

TABLE 25 (continued)

Period	Per Cent Removal (%)				F/M	Mixed Liquor Temp
	BOD_5 PEUF-FEF	COD PEUF-FEF	TSS	VSS	$= \frac{\text{lb. BOD}}{\text{day, lb MLVSS}}$	^{o}C
January 1 to January 31	89.6	84.7	82.0	82.5	0.26	≈ 14
February 1 to February 12	92.1	83.7	77.4	72.8	0.19	≈ 13
February 13 to February 17	90.4	84.5	87.4	85.6	0.24	≈ 13
March 12 to March 17	92.2	78.8	91.0	88.7	0.21	≈ 11
March 26 to April 5	89.7	79.1	73.0	71.2	0.33	≈ 12
April 6 to April 22	91.8	79.1	71.4	70.2	0.51	≈ 14
April 24 to May 9	91.0	80.5	83.2	81.7	1.00	≈ 15

Parallel Operation

June 15-30:

Table 15 clearly indicates that the detention time and the aeration tank solids concentration were maintained relatively the same in both plants. BOD and solids removals were equally efficient. There was a slightly higher COD removal in the full scale plant, but there was data for only four days for the pilot plant. The slightly lower F/M ratio, i.e., lbs BOD/day/lb MLVSS in the pilot plant, is due to a lower measured BOD influent (sampling points for model and prototype are located at different places).

July 1-11 and 26-31:

Referring to Table 16, the comparisons of loadings and removal efficiencies are fairly good in most cases. The most notable exception is the suspended and volatile suspended solids removal. This is attributed to operational problems in the pilot plant arising out of malfunctioning of the sludge scraper at the bottom of the final settling tank.

July 12-25:

In this period too, most parameters are well matched. For reasons mentioned above, the solids removal efficiency in the pilot plant is considerably lower.

August 10-19:

This period, summarized in Table 18, is undoubtedly the best period as far as soluble BOD removal efficiency is concerned. There is slightly lower COD removal in the pilot plant. Suspended solids and volatile suspended solids removal is again significantly higher in the full scale plant. Referring to the aeration tank data for this period, it is seen that a fairly high DO was maintained during this period giving an extremely high value of air applied per lb of BOD. This might have been one cause of solids carry-over in the final settling tank.

In summarizing the parallel operation phase, it is seen that in most cases removal efficiencies and other parameters matched quite closely. One parameter that was significantly different in each case was the loading on the final settling tanks; both hydraulic and solids loading being much higher in the full scale plant.

For the constant T operation, it must be mentioned again that the pilot plant was operated separately from the full scale, i.e., at the time of constant T operation in the pilot plant there was no control over the flow in the full scale. The comparisons made are with similar full scale operation at an earlier time. The only common factor is similar nominal detention time. There was no attempt to match mixed liquor solids concentrations.

Constant Detention Time Operation

5 Hour Nominal T:

Table 19 shows that there is little difference in average removal efficiencies, especially for soluble BOD and COD values. Influent BOD values are quite different, but owing to different MLSS concentrations, the F/M ratio is comparable.

4 Hour Nominal T:

As can be seen in Table 20, most of the removal efficiencies and parameters are quite comparable. It is interesting to note a higher soluble BOD removal in the pilot plant even though MLSS concentrations is nearly half of that in the full scale.

3 Hour Nominal T:

This period shows an excellant comparison between the two systems, with the pilot plant having a slight edge over the full scale.

2 Hour Nominal T:

Soluble BOD removal efficiency increased in both plants over the 3 hour period. Similar trends in both the plants is quite significant.

1 Hour Nominal T:

The pilot plant showed a slight decrease in BOD removal efficiency. Solids removal efficiency is significantly lower. Table 24 and 25 summarize constant T operation for pilot and full scale plants respectively. One marked difference is that the mixed liquor temperatures were higher for the pilot plant than for the full scale.

Oxygen Transfer Tests

Two series of oxygen transfer tests (using the re-oxygenation technique[12,13])were run on the pilot plant utilizing tap water and final effluent from the full scale plant. A summary is given in Table 26. Transfer efficiency decreases with increase in air rate as shown in Figure 66.

24 Hour Grab Evaluation

Figures 67-69 show the results of an intensive 24 hour evaluation where sampling was done every two hours and DO was maintained constant. Figure 67 is the data for variation in primary effluent parameters. Figure 68 is a similar plot for the final effluent, and Figure 69 gives the aeration tank DO and air supplied over the same period. It is seen that the peak values for each parameter came at 12 noon. An initial peak in some cases at 12 mid-night is because a little before that time, supernatant from the digesters was pumped through the primary settling tanks. It is interesting to note that the maximum air was supplied immediately after 12 noon. The total air supplied in the 24 hour period was estimated to be 270 scf, giving an average DO of 2.1 mg/l.

Tracer Test

The washout curve obtained in the tracer test is shown in Figure 70. The data points show a logarithmic decay curve typical of a completely mixed system. The peak concentration was reached in just under ten minutes. A K of 0.01007 min^{-1} was measured as the slope of a semi-log plot of C_i/C_o versus t_i/T, giving a measured T value of 1.66 hours as compared with a T theoretical of 2.05 hours. This indicates that the effective tank volume was somewhat less than the actual. n was calculated to be 1.1, which is good compared to an ideal complete mix value of 1.0. Dye recovery was calculated to be 89%.

Oxygen Uptake Tests

As in the full scale studies, oxygen uptake tests were run on the pilot plant mixed liquor. A typical test result is plotted in Figure 71. The tests and measured and calculated values have been summarized in Table 27. An $\frac{M_a}{VSS}$ ratio of 0.26 was obtained; the average values for K_2 and K_9 being .029 and 1.44 hrs^{-1} (O_2 Units) respectively (K_2 is equivalent to K_e and K_9 is equivalent to K_s).

SUMMARY OF OXYGEN TRANSFER TESTS

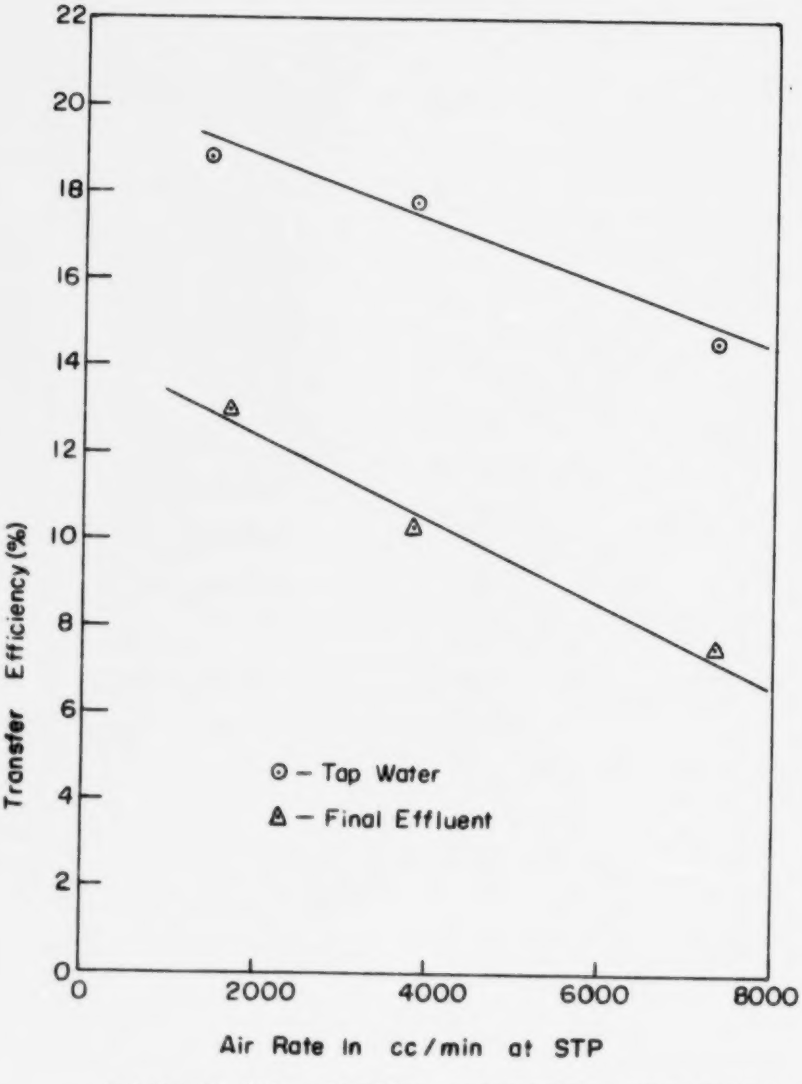
Table 26

TAP WATER

Run No.	1	2	3
Air Rate in cc/min	1450	3850	7350
K20 min^{-1}	.0803	.202	.314
Transfer Efficiency %	18.8	17.8	14.6
Liquid Temp. °C	16.7	16.7	16.7

FINAL EFFLUENT

Run No.	1	2	3
Air Rate in cc/min.	1700	3850	7350
K20, min^{-1}	.0638	.119	.167
Transfer Efficiency	13.0	10.3	7.6
Liquid Temp. °C	19.0	19.0	19.0
$\alpha = \frac{KFE}{K \text{ tap water}}$	.79	.59	.53



PILOT PLANT OXYGEN TRANSFER EFFICENCY
Figure 66

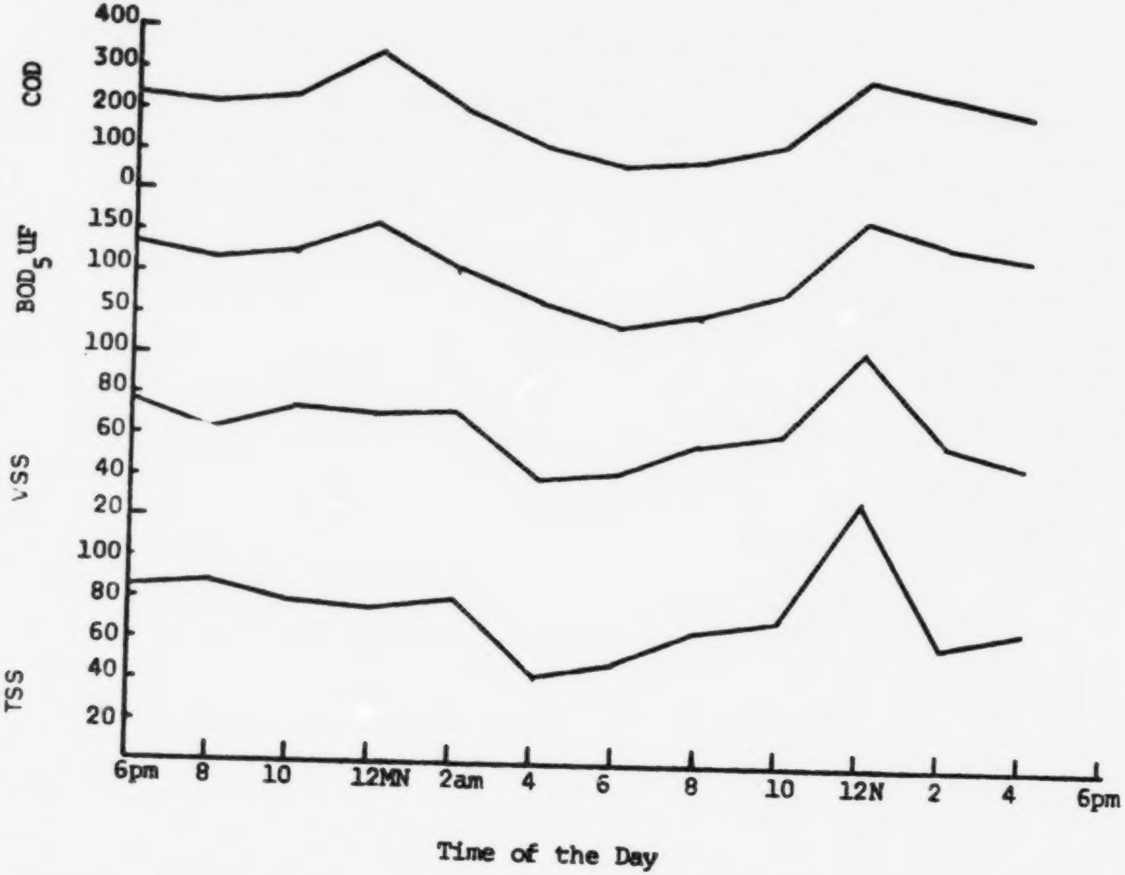


Figure 67 24 Hour Grab Analysis - Primary Effluent

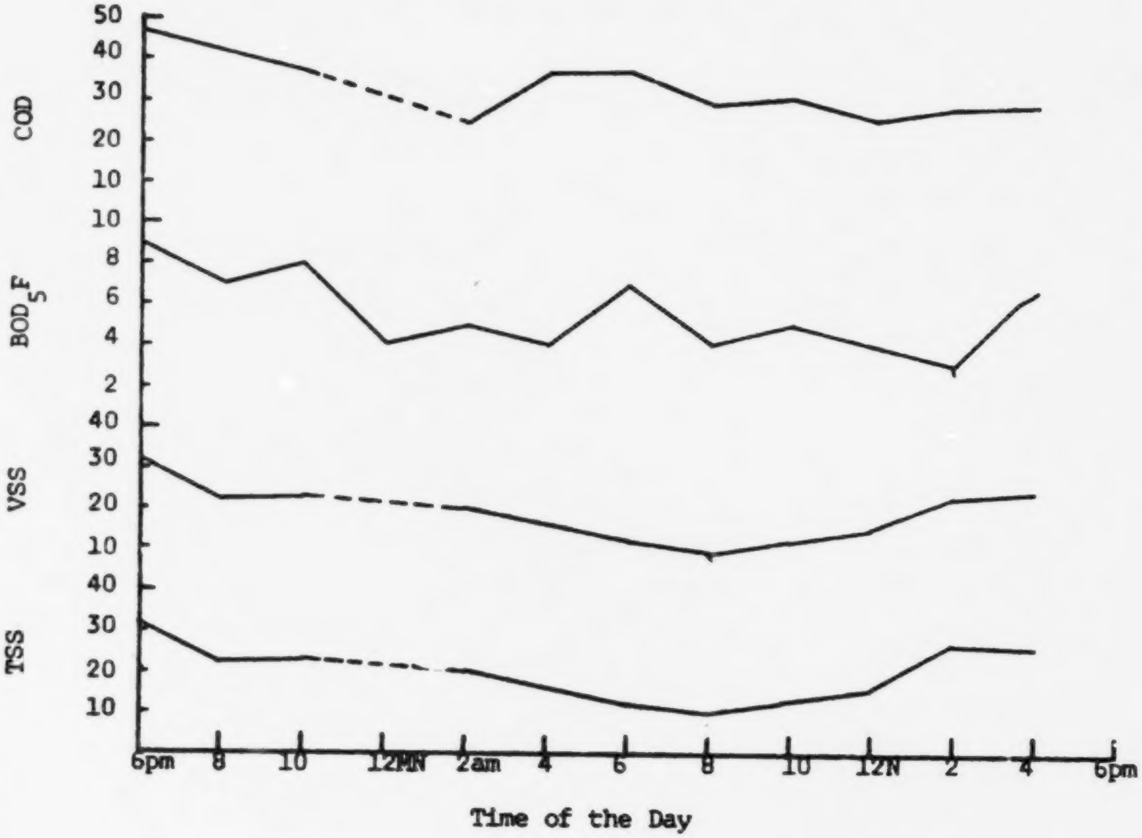


Figure 68 24 Hour Grab Analysis - Final Effluent

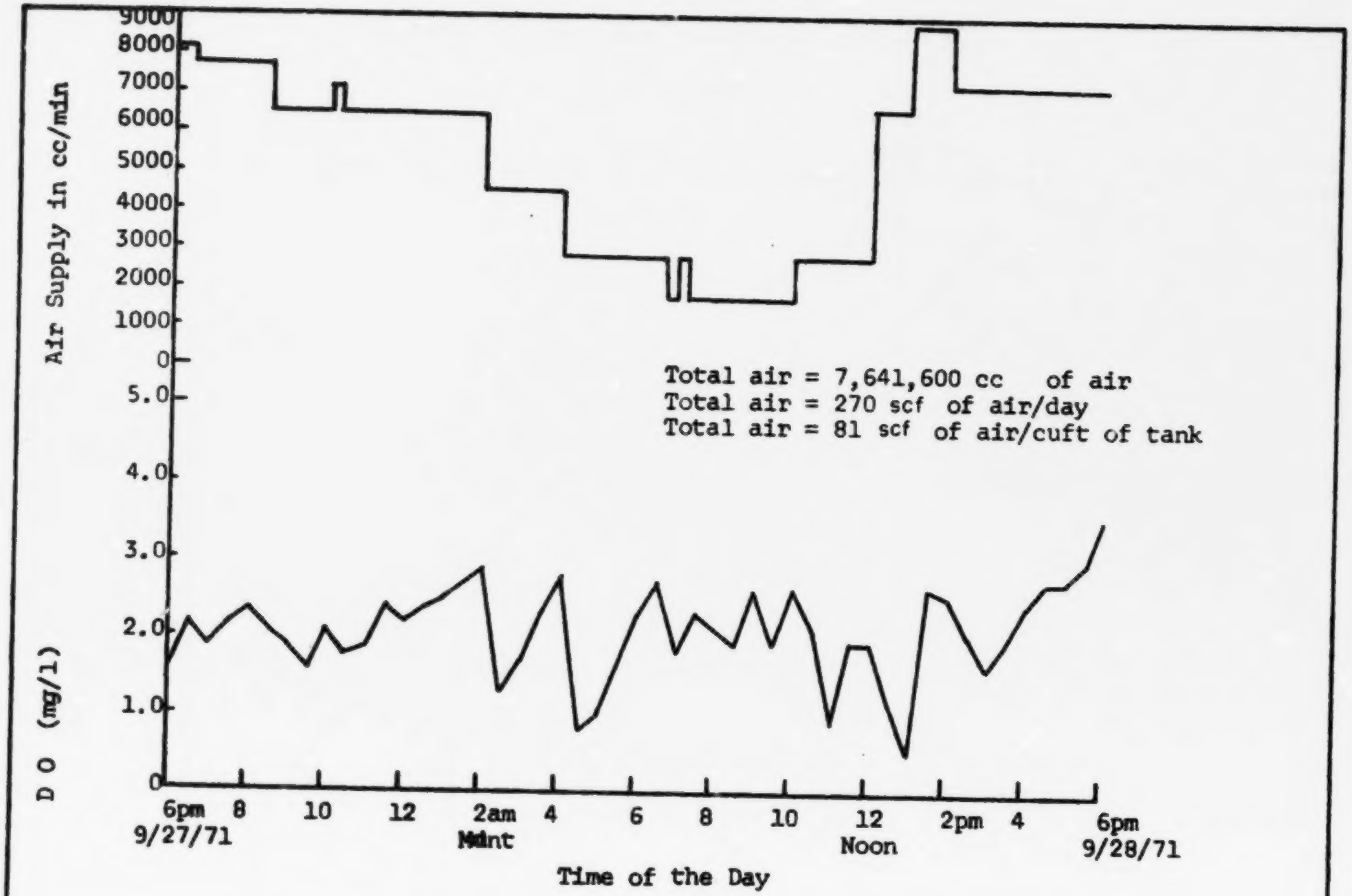


Figure 69 24 Hour Evaluation - DO profile and air supplied

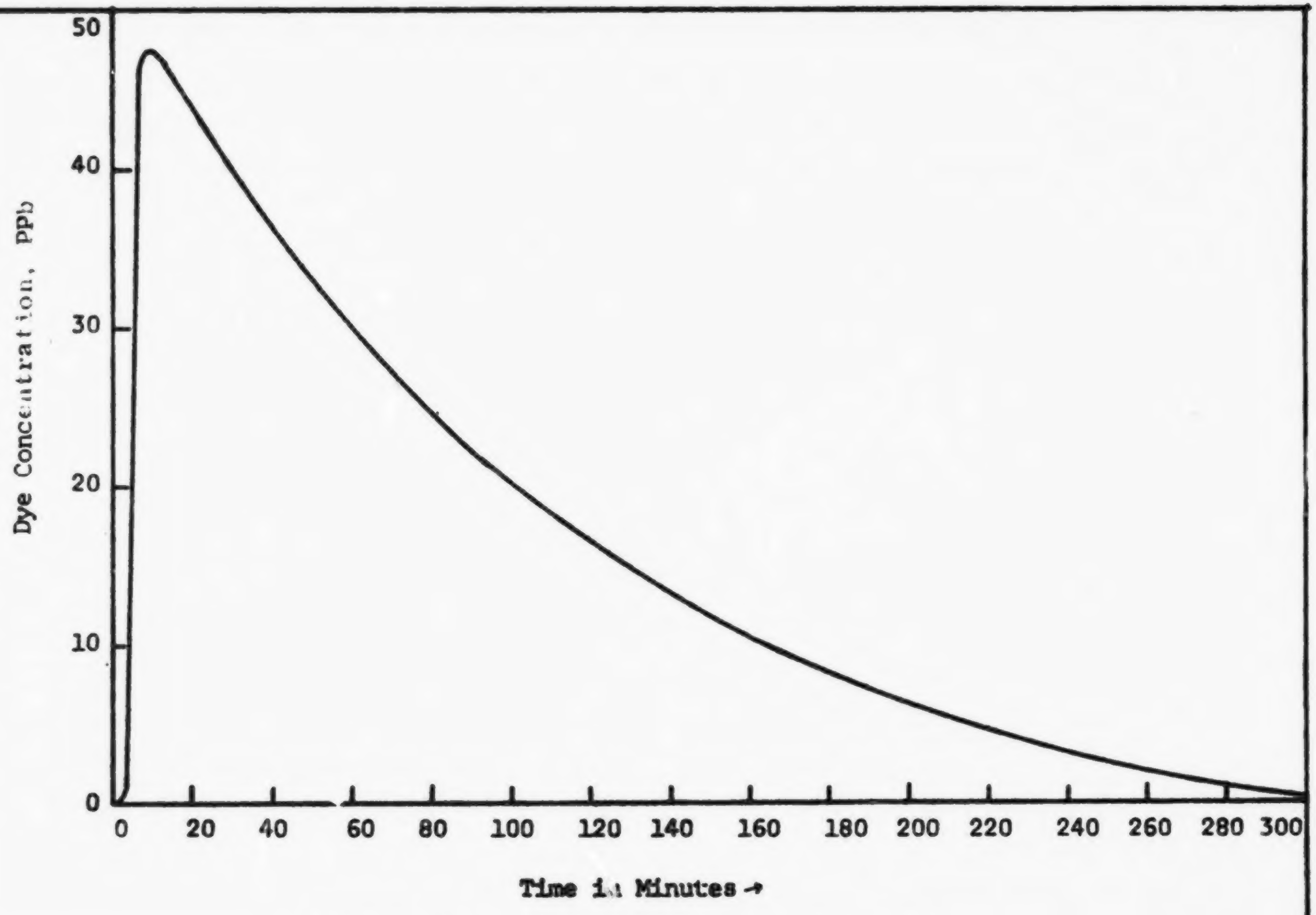


Figure 70 Dye Wash Out Curve, Tracer Test

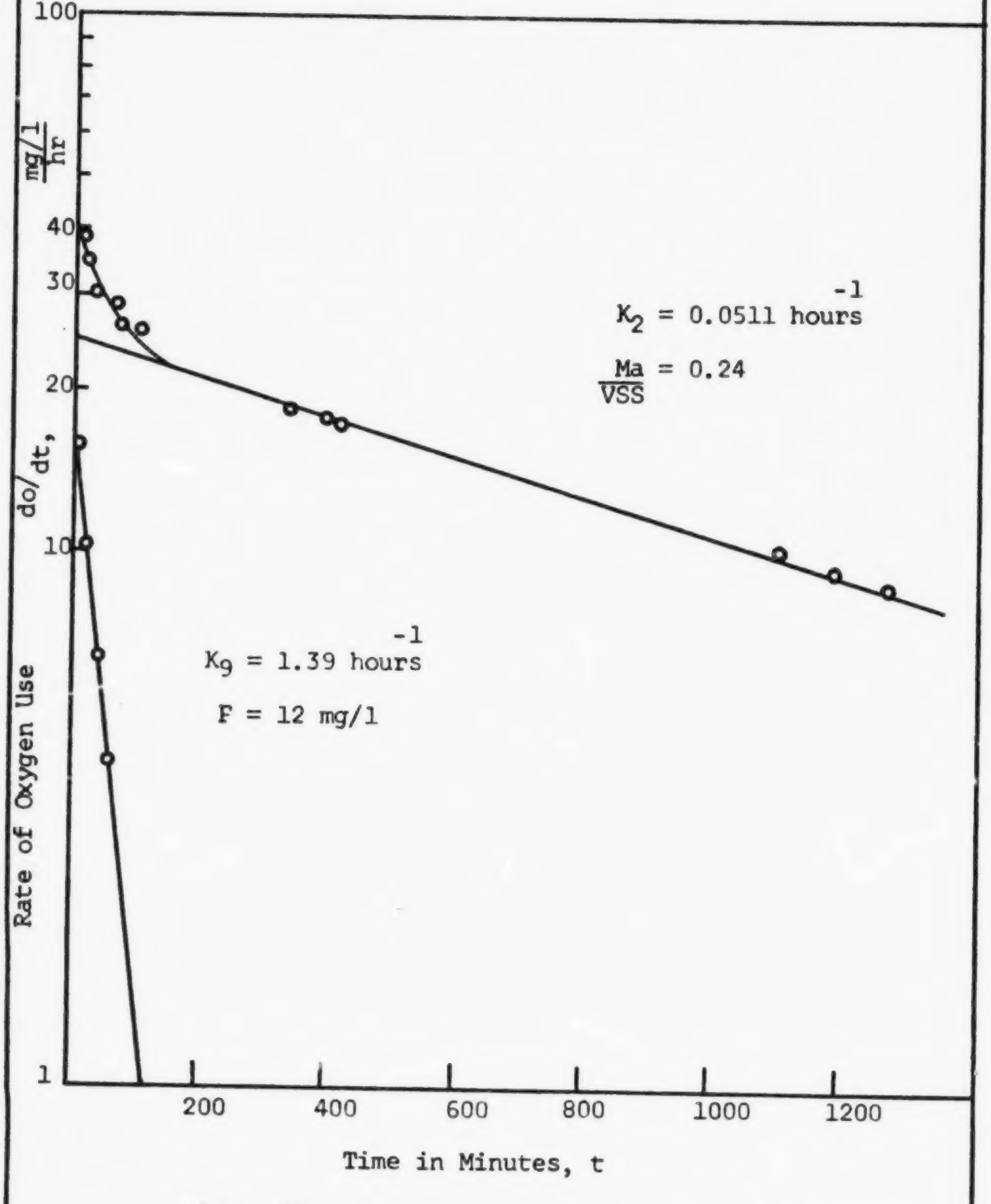


Figure 71 Oxygen Uptake, Run No. 9/20/99

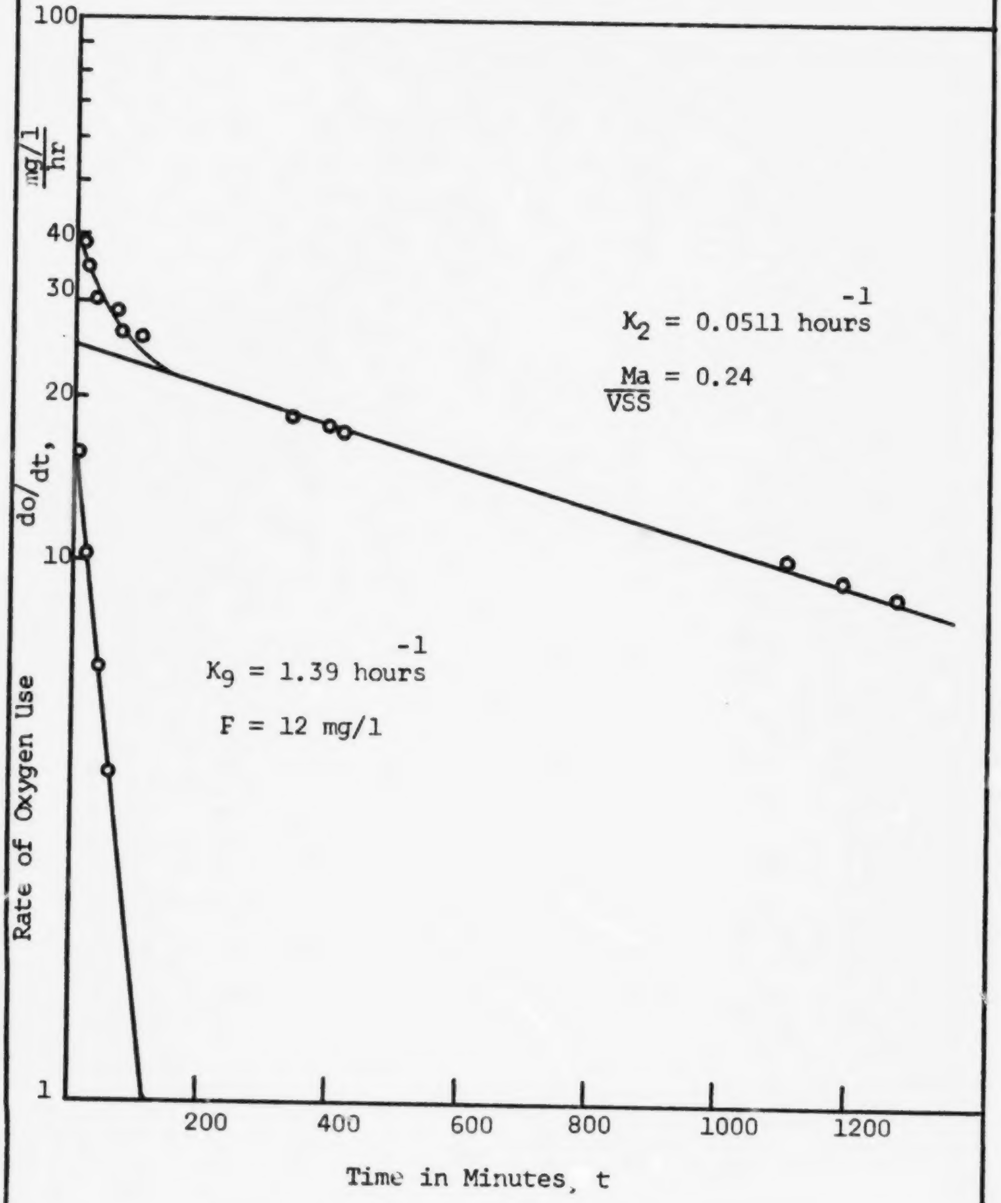


Figure 71 Oxygen Uptake, Run No. 9/20/99

TABLE 27

Summary of Oxygen Uptake Tests

Date	Sample	K_2, hrs.$^{-1}$	K_9, hrs.$^{-1}$	F, mg/l	Ma, mg/l		VSS, mg/l	$\frac{Ma}{VSS}$	Oxygen Uptake in mg/l/hr due to		Total Uptake mg/l/hr	Total Uptake mg/l/day
					O_2	Solids			Synthesis	Endogenous		
7/1	M.L.	.0296	2.67	3.0	450	317	1190	0.26	7.7	13.5	21.0	504
8/14	M.L.	0214	1.19	5.0	467	329	1210	0.27	6.0	10.0	16.0	384
8/31	M.L.	.0262	1.73	2.0	382	269	1330	0.20	3.0	10.0	13.0	312
9/9	M.L.	.0269	0.59	17.0	409	288	2140	0.13	10.0	11.0	21	504
9/20	M.L.	.0511	1.39	12.0	489	345	1450	0.24	16.0	25.0	41	984
9/29	M.L.	.0199	1.04	12.0	980	690	1570	0.44	12.0	19.5	31.5	756
Average:		.029	1.44					0.26				
Average from Full Scale operation:		.026	1.19					0.26				

* M.L. = Mixed Liquor

Discussion

The following points should first be noted:

1. The results of the tracer test clearly indicate that the system operated under nearly ideal complete mix conditions.
2. For all computations and analyses, except removal efficiencies, the BOD values used are ultimate values. These are based on the relation, $\frac{BOD_L}{BOD_5} = 1.46$. The

BOD rate constant was determined experimentally for full-scale samples and was calculated to be 0.1 for both the primary and the final effluent samples. The same value was accepted for the pilot plant because of very similar sample properties.

Referring to the parallel and constant detention time operations (Tables 15 to 25), the following observations can be made. In the periods of parallel operation, when flows were duplicated and mixed liquor solids levels were matched, removal efficiencies and other parameters showed good comparison between the pilot and full-scale plants.

The highest removal efficiency (based on soluble BOD_5) predicted by the pilot plant was 91% and the highest achieved in the full scale plant was 93%- the two values falling in different periods of parallel operation. During the constant detention operation, the highest removal efficiencies were 94% and 92% respectively for the model and prototype.

It should be born in mind that all removal efficiencies were based only on primary to final effluent. If they had been based on raw to final, higher efficiencies would obviously have resulted. The above discussion on removal efficiencies thus restricts itself to the activated sludge process only.

In Figures 67-69, the data from 24-hour grab analysis has been plotted. Referring to Figure 67, the plot for the primary effluent, it is seen that the strength of the waste changes considerably during the day; the BOD varying from a high of 150 mg/l to a low of 50 mg/l. Thus, there is an appreciable fluctuation in oxygen demand as is well defined in Figure 69. It is therefore quite possible that due to oxygen becoming limiting, complete biological oxidation is lacking, resulting in high final effluent BOD values. Figure 68 indicates that if enough air is applied at all times, a consistently low final effluent BOD can be obtained.

The COD measurements do not seem to show much variation from period to period, though at high detention times there is positive evidence of better removal.

McKinney's BOD removal rate constant, K_4, was determined using data from each period. The results have been summarized in Table 28 for the pilot plant operation. The values of K_4 range from 2.4 hrs^{-1} to 7.3 hrs^{-1}. Table 29 compares removal rate constants for four periods of full scale data. K_4 values from both plants, especially for BOD_L, are extremely comparable with both averaging around 2.9 hrs^{-1}.

Figure 72 is a plot of % BOD removal versus detention time T. The solid line has been drawn for the constant T operation. The dashed lines are theoretical curves for K_4 varying from 2 to 8 hrs^{-1}. It is seen that the experimental curve falls between $K = 2$ and $K = 5$. This value of K_4 is considerably lower than those reported by McKinney and Wallace and others. McKinney reported values in the range of 15 hrs^{-1}.

Though McKinney[10] and Smith[7] advocate no dependency of K_4 on MLSS, Wallace[15] in his work at the University of Iowa showed definite dependency of K_4 on MLSS in some constant T periods and a lack of any correlation in others. Figures 73-76 were plotted (for two periods) and indicate a relationship between K_4 and MLSS. In the present investigation at nominal T values of 5 hours and 3 hours, there is every reason to believe that K_4 is solids dependent. There does not seem to be any correlation in any of the other time periods. One thing, however, must be born in mind here. In all periods except one, the maximum level of MLSS concentration is about 3000 mg/1.

That the removal rate constant is solids dependent was proposed by Eckenfelder[11] in his model. Data for this analysis is plotted in Figures 77 and 78. There is a trend towards a linear relationship, and a straight line approximated through the data points gives a K of 0.064 $days^{-1}$ for BOD and .031 $days^{-1}$ for COD. If we were to compare the removal rate constants for the two different theories we would have McKinney K_4 = Eckenfelder K x MLSS. With respect to the BOD analysis then, for the whole pilot plant operation Eck $K = .064$ $days^{-1}$ and average $K_4 = 90.6$ $days^{-1}$. For Eck K the equivalent K_4 would then be K x MLSS (avg) $= 103.3$ $days^{-1}$, which gives a 12% difference.

Another attempt was made to show the extent of correlation between K and K_4. If we assume an unfiltered $PEBOD_L$ of 100 mg/1 and a $FEBOD_L$ of 10 mg/1, using an average K_4 of 3.8 hrs^{-1} we have t as 2.4 hrs = 0.1 days. If we now go to Figure 77, for $S_e = 10$ mg/1 we have $\frac{S_o - S_e}{X_a\, t} = 0.55$ from

TABLE 28

BOD Removal Rate Constants

	K_4*	
Period	days^{-1}	hours^{-1}
1	58.7	2.4
2	57.9	2.4
3	66.0	2.8
4	90.0	3.8
5	78.4	3.3
6	47.2	1.9
7	88.0	3.7
8	155.0	6.5
9	174	7.3

* $$\frac{F}{F_o} = \frac{1}{K_4 t + 1}$$

F_o = BOD L PEUF

F = BOD L FEF

TABLE 22

Comparison of K_4 Values for Pilot Plant and Full Scale

Period	K_4, days^{-1} BODL Basis P	BODL Basis F	COD Basis P	COD Basis F
1	58.7	47.4	25.4	39.8
2	57.9	60.8	22.9	22.2
3	66.0	98.5	26.9	34.3
4	90.0	74.2	27.9	43.2
Average	68.2	70.2	25.8	34.9

Period	K_4, hours^{-1} BODL Basis P	BODL Basis F	COD Basis P	COD Basis F
1	2.4	1.9	1.1	1.7
2	2.4	2.5	0.9	0.9
3	2.8	4.1	1.1	1.4
4	3.8	3.1	1.2	1.8
Average	2.85	2.90	1.1	1.5

P: Pilot
F: Full-Scale

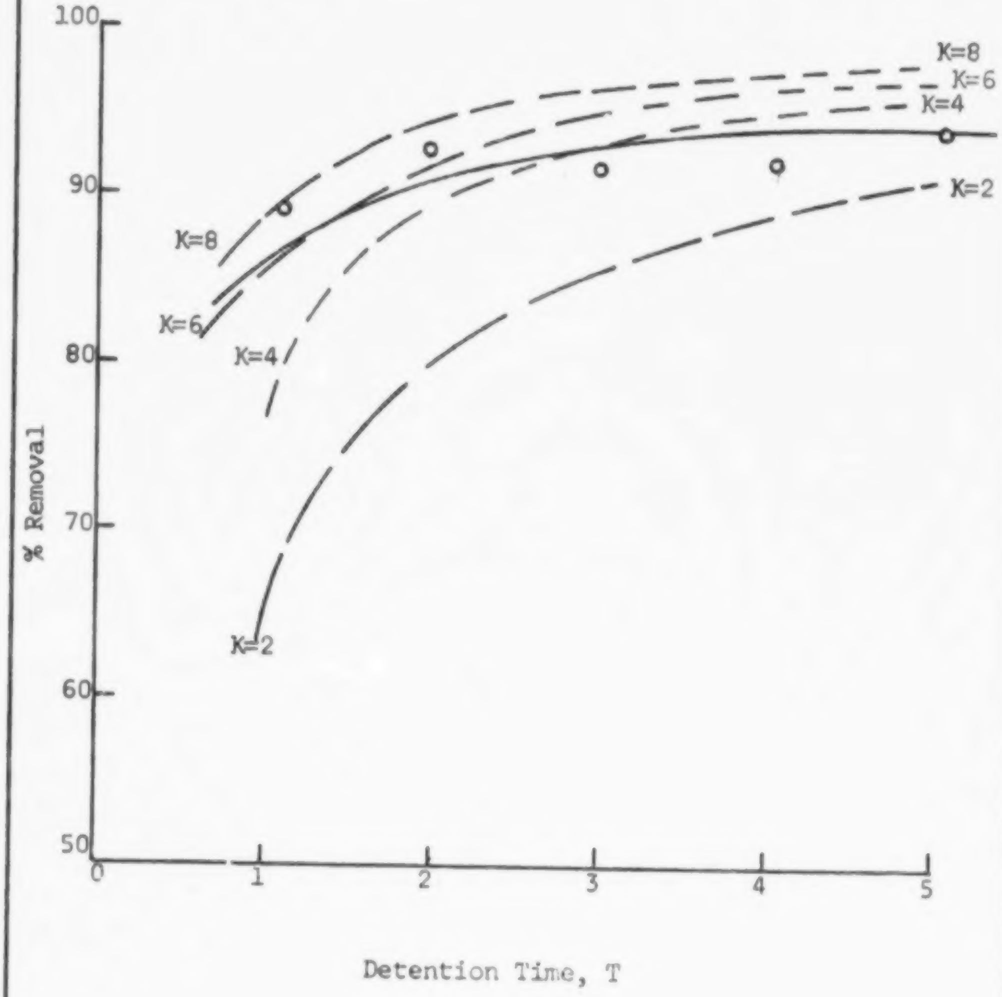


Figure 72 % BOD Removal Versus Detention Time, Constant T Operation

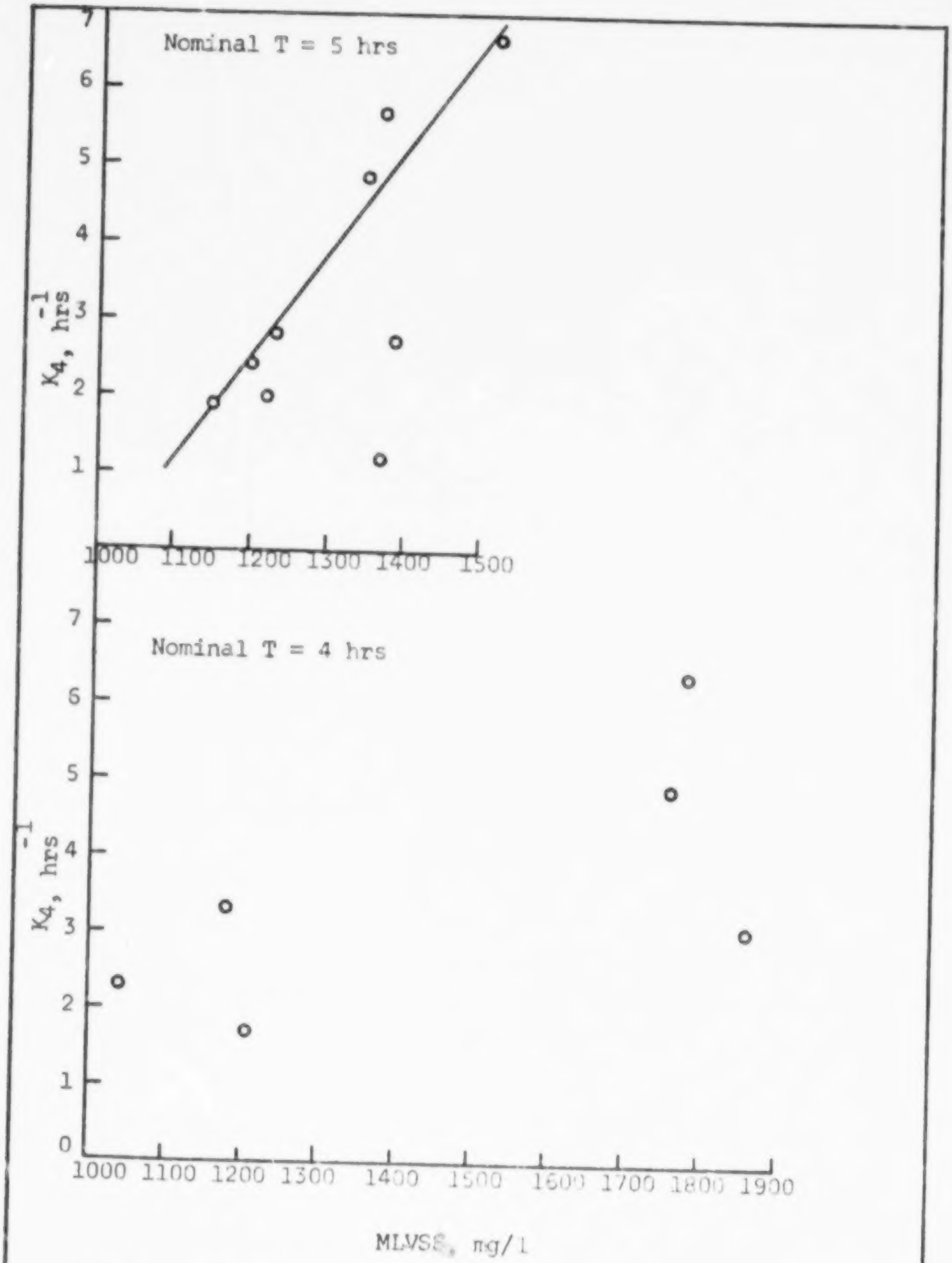


Figure 73 K_4 versus MLVSS, 5 and 4 hrs. Nominal T

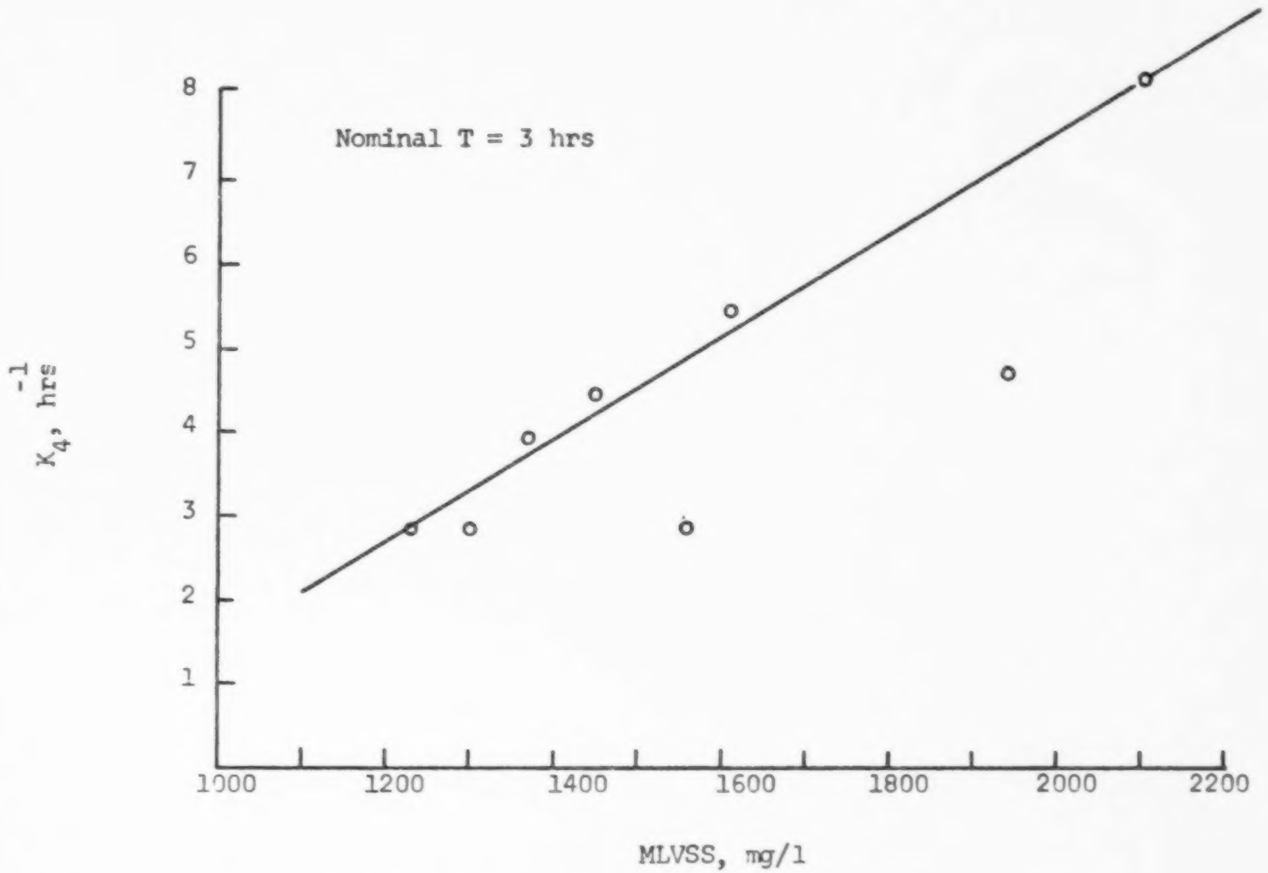


Figure 74 K_4 versus MLVSS, 3 hrs Nominal T

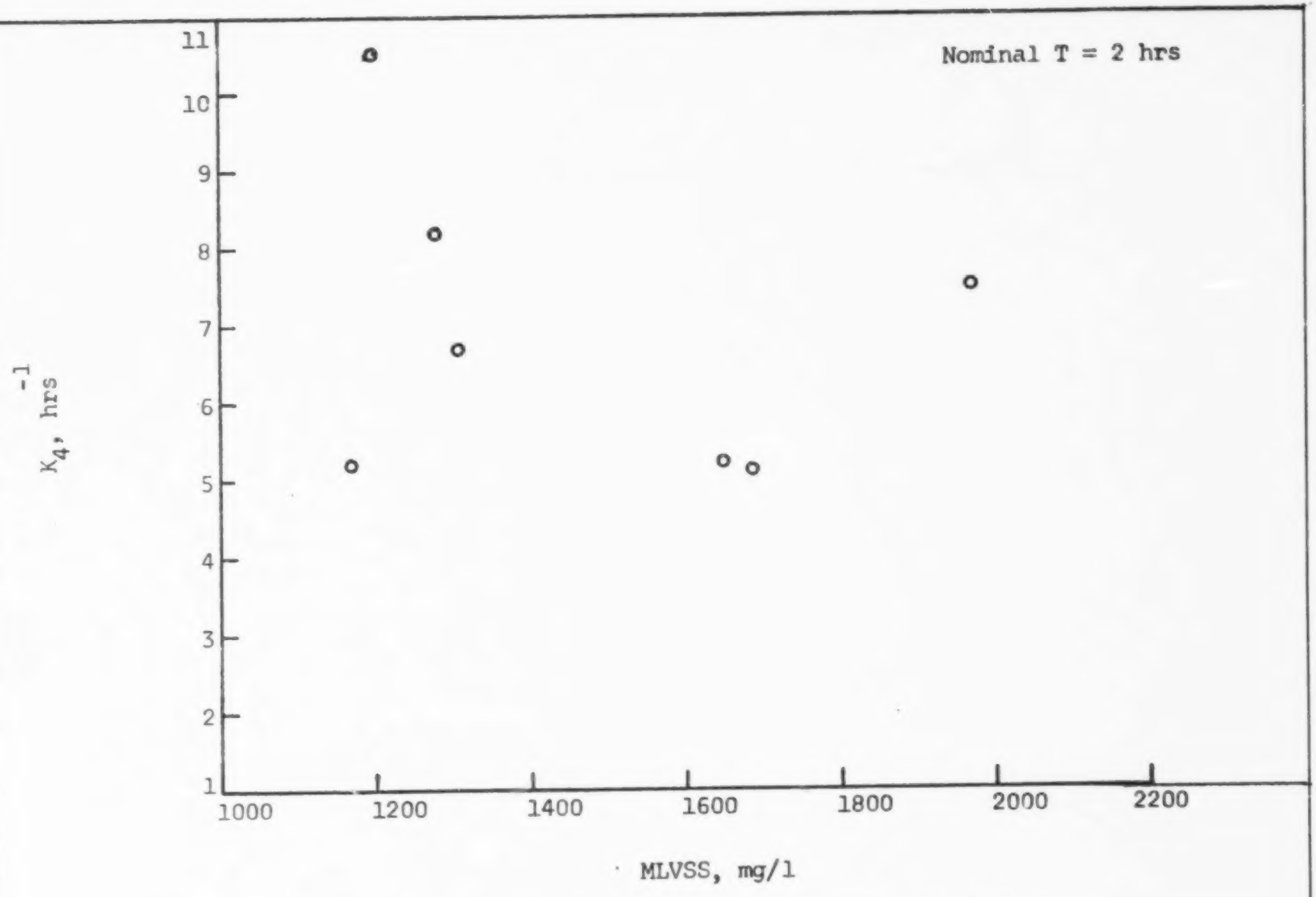


Figure 75 K_4 versus MLVSS, 2 hrs Nominal T

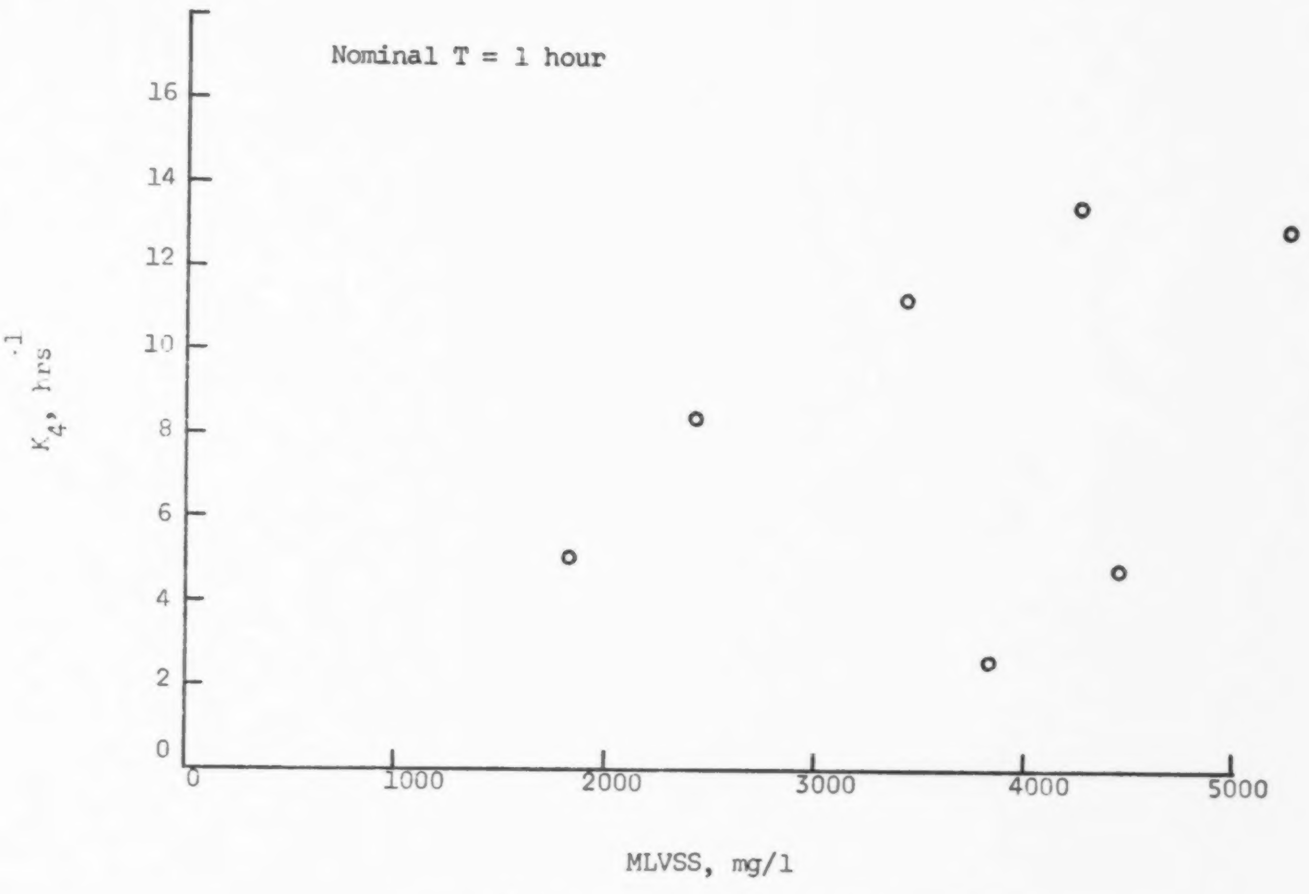


Figure 76 K_4 versus MLVSS, 1 hr Nominal T

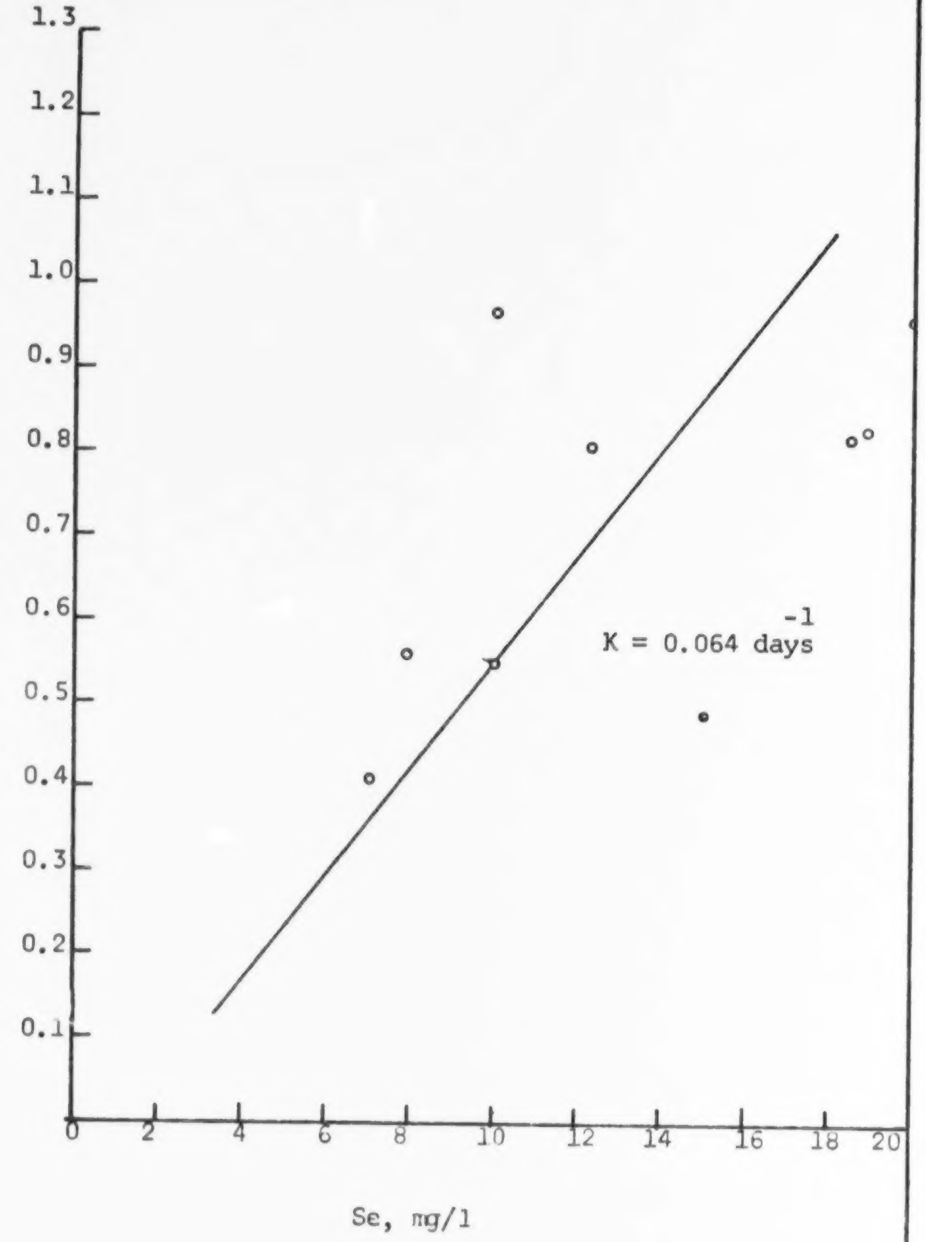


Figure 77 Removal Rate Constant, K, Eckenfelder Approach - BOD_L Basis

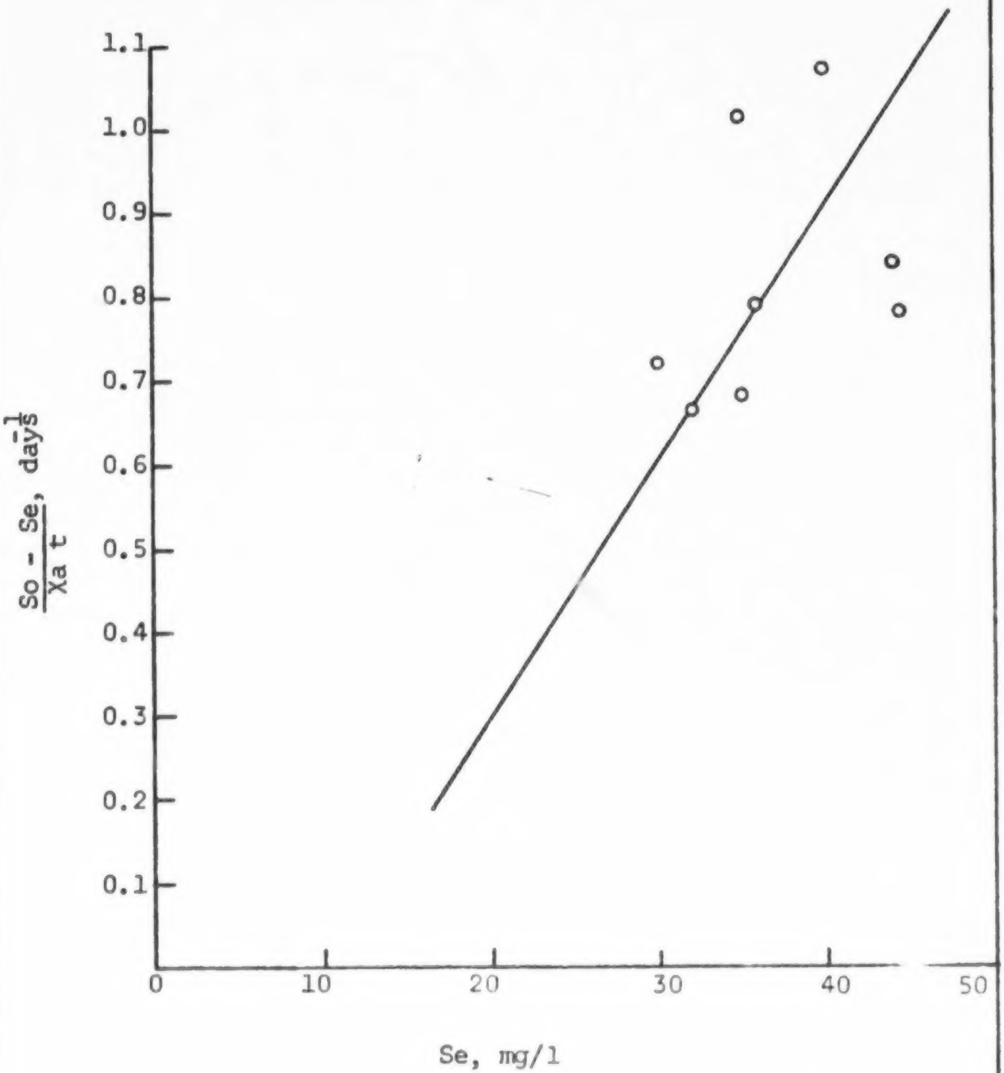


Figure 78 Removal Rate Constant
Eckenfelder Approach - COD Basis

which value an X_a of 1640 mg/l can be determined. Then $K_4 = K \times MLSS = \frac{.064 \times 1640}{24} = 4.3\ hrs^{-1}$ which compares well with a measured K_4 of 3.8 hrs. K and K_4 do not necessarily yield the same detention time for a given set of S_o, S_e and X_a values. Data available is inadequate to make definite conclusions about the relative preference of one constant over the other.

In continuing the analysis of the operating data, consider Table 27 which summarizes O_2 uptake data. $K_2(K_e)$ and $K_9(K_s)$ were averaged out to be 0.029 hrs^{-1}, O_2 units (.041, solid units) and 1.44 hrs^{-1}, O_2 units, respectively, and average $\frac{M_a}{VSS}$ was 0.26, which compares extemely well with an $\frac{M_a}{VSS}$ of 0.26 on the full scale and a value of 0.27 predicted by Frazier[16].

In the oxygen uptake analysis, determination of K_2 is more reliable than the corresponding K_9 value which depends on initial uptake rates which change quite rapidly. Thus, accepting an average value for K_2, values for K_9 were determined from the operating data (using the equation $M_{ae} + wM_{ar} = T(K_sF - K_eM_a)$ for each period of pilot plant operation These have been given in Table 30. Since the oxygen uptake tests for the pilot plant were conducted using only mixed liquor samples, values for $\frac{M_{ar}}{VSS}$ and $\frac{M_{ae}}{VSS}$ were assumed as 0.26 and 0.4 respectively, equal to those from full scale data. K_9 was averaged to be 2.1 hrs^{-1}, neglecting values greater than 3 hrs^{-1}.

Values of 0.029 hrs^{-1} for K_2 and 2.1 hrs^{-1} for K_9 were then used in predicting oxygen uptakes from the mass balance equation and consequently used to predict air requirements.

Referring back to Figure 69 for the 24 hour period when an average DO of 2.1 mg/l was maintained, about 270 scf of air was supplied. The following is a theoretical air requirement analysis based on K_2, K_9 and the active mass equation.

From the oxygen uptake data and the operational data, we have K_2 and K_9 as 0.041 hrs^{-1}, solids units (.029 hrs^{-1}, O_2 units) and 2.1 hrs^{-1}, O_2 units, respectively, and $\frac{M_a}{VSS}$ is calculated to be 0.26. For this period of the 24 hour grab evaluation, the MLVSS was averaged out to be 2655 mg/l, and the final effluent filtered BOD_L was estimated to be 12 mg/l. The total uptake is then given by $\frac{do}{dt} = K_9F +$

TABLE 39

Computed Values for K_9

Period	* K_9 in hours
1	1.53
2	1.25
3	1.86
4	3.46
5	2.84
6	1.32
7	2.78
8	2.84
9	4.13

Neglecting values greater than 3, Avg. $K_9 = 2.1$ hours.

Average from full scale operation $K_9 = 2.4$ hours.

*Based on:

$$Mae + w\ Mar = T(K_9F - K_2Ma)$$

K_2M_a, which gives in terms of oxygen units a total requirement of 1287 mg/l/day. The 270 scf/day of air supplied can be converted into similar units to facilitate a comparison. An oxygen transfer efficiency of 9.5% is estimated as follows. The total air supplied was 7,641,600 cc/day (Figure 69). This can be converted to 5307 cc/min. Knowing the air rate, the transfer efficiency is obtained from the final effluent curve in Figure 66. The oxygen transfer rate can then be calculated as follows: $\frac{do}{dt} = \frac{270 \text{ scf}}{\text{day}} \times$

$$\frac{.0174 \text{ lb } O_2}{\text{scf}} \times \frac{.095 \times 453.6 \times 10^3 \text{mg}}{94.62 \text{ liters} \quad \text{lb}} = 2140 \text{ mg/l/day}$$

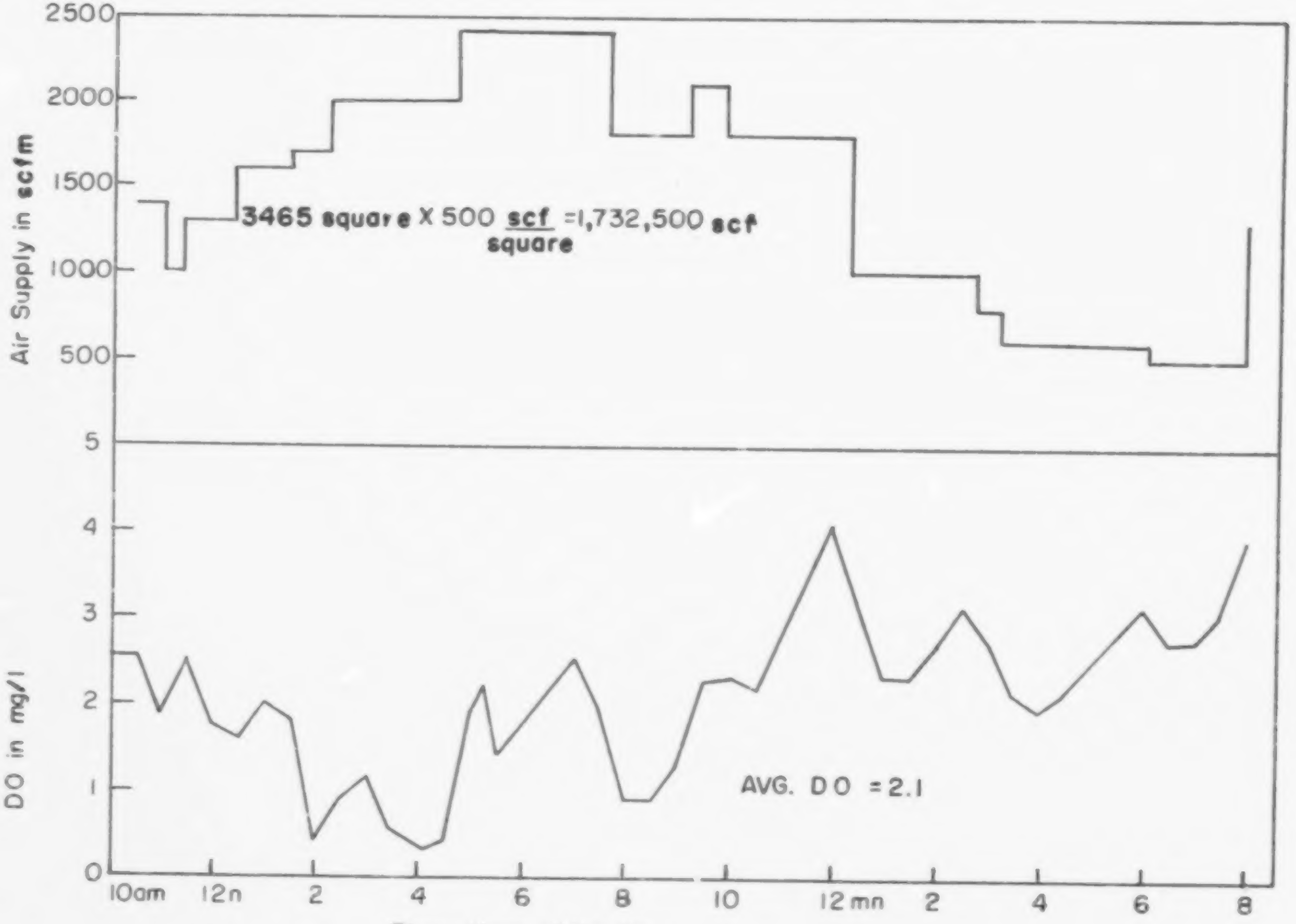
This indicates that more air was supplied than predicted by the uptake equation above.

To study the capacity of the pilot plant to predict air requirements for the full scale plant, a similar analysis was performed using the full scale operating data. Figure 79 is a plot of a 24 hour DO monitoring of the full scale plant. The total air supplied was estimated at about 1.7 million scf on September 8-9. This gives an air rate of about 600 scfm per tank. In the same period, the following values were obtained from the full scale operation:

$$\text{MLVSS} = 2113 \text{ mg/l}$$

$$\text{BOD}_L\text{, FEF} = 10 \text{ mg/l}$$

The total uptake is then $do/dt = K_9F + K_2M_a = 1043$ mg/l/day. Air supplied to the full scale tank can be converted to the same units as before. From oxygen transfer efficiency tests on the full scale aeration tank, the transfer efficiency was estimated to be 10% at an air rate of 500 scfm per tank(Note the similarity in the pilot and full scale oxygen transfer efficiencies). Using a transfer efficiency of 10%, the rate at which air was supplied is calculated to be 1370 mg/l/day, which does agree closely with the 1046 mg/l/day required, keeping in mind that DO's did fluctuate above 2 mg/l during the 24 hour period.



DO PROFILE, SEPT. 8-9 1971, FULL SCALE PLANT
Figure 79

Solids Production

The two methods of solids production analysis (active mass balance and Eckenfelder solids balance) outlined in Section V were applied to the pilot plant data as they were to the full scale data.

The following is an analysis of solids production using the Eckenfelder approach.

Figure 80 is a plot of $\frac{\Delta X}{X_a V}$ versus $\frac{S_r}{X_a \cdot t}$ from the data in Table 31. The constants a and b are calculated as 1.0 (dimensionless) and $^-0.22$ (days^{-1}) respectively. The dashed line is a similar plot for full scale operations which gives a and b as 0.8 and -0.10 respectively. The final prediction equation for the pilot plant is then,

$$\frac{\Delta X}{X_a V} = 1.0 \frac{S_r}{X_a t} - 0.22$$

It is interesting to note that the a and b values for the pilot plant and the full scale PF solids analysis agree exactly (a = 1, b = -.22)

Solids production was also estimated using the basic active mass equation, $M_{ae} + wM_{ar} = T(K_s F - K_e M_a)$ from which w can be determined since all other constants have already been established.

Then sludge wasted is given by $\Delta X = wQ \times RSVSS$.

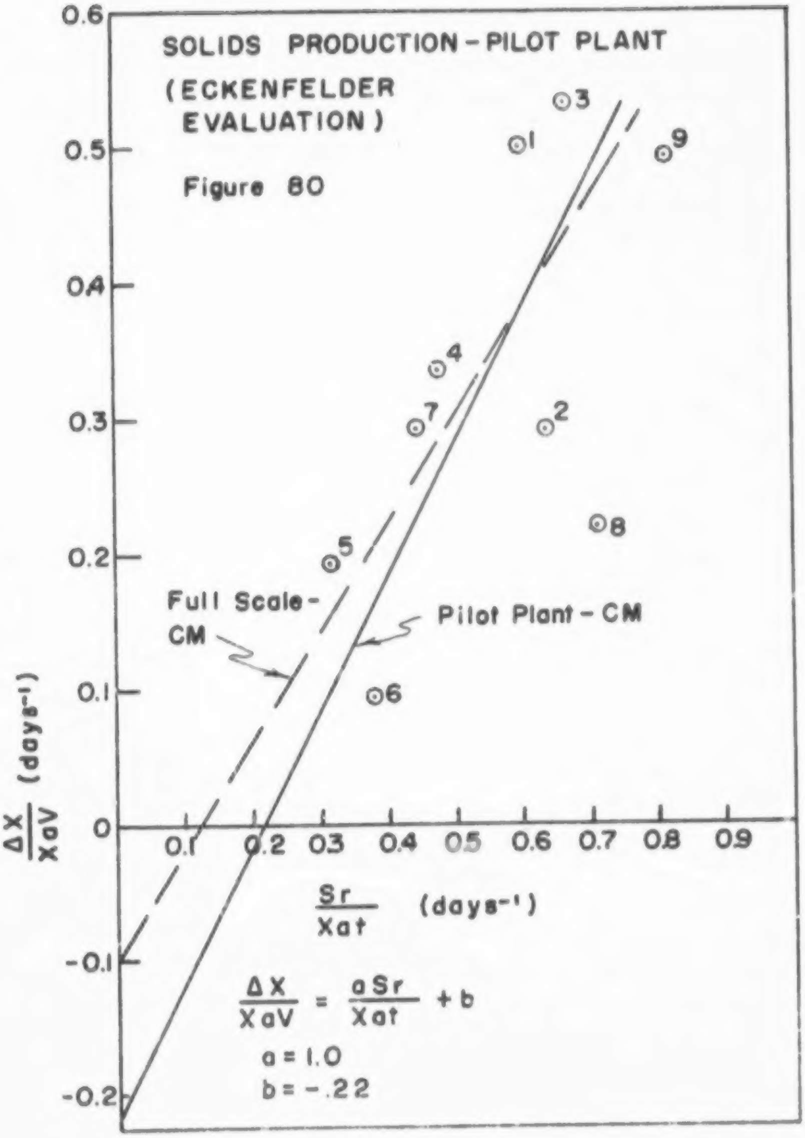
To study the applicability of each model and to see how well the pilot plant constants could predict sludge accumulation, values of ΔX were calculated from each model using the data from the full scale CM operation. These predicted values of ΔX were then compared with the values of sludge wasted actually recorded during full scale operations. Tables 32 and 33 respectively show measured and predicted values for sludge wasted using the active mass evaluation and the Eckenfelder evaluation.

To evaluate the Eckenfelder approach, $\frac{S_r}{X_a \cdot t}$ was calculated for each full scale operational period. The corresponding $\frac{\Delta X}{X_a V}$ was estimated from Figure 80 and ΔX calculated from operational values of X_a and V. For the active mass balance model, actual values were used in the active mass equation to calculate w and finally ΔX. In this analysis

TABLE 31

Sludge Production - Eckenfelder's Evaluation

No.	Period	$\frac{\Delta X}{X_a V}$ (days^{-1})	$\frac{S_r}{X_a t}$ (days^{-1})
1	June 15-30, 1971	0.50	0.60
2	July 1-11 & 26-31	0.29	0.63
3	July 12-25	0.53	0.67
4	August 10-19	0.33	0.48
5	August 20-31	0.19	0.32
6	Sept. 1-10	0.09	0.38
7	Sept. 11-20	0.29	0.45
8	Sept. 21-30	0.22	0.71
9	April 27 - May 7	0.49	0.82



SOLIDS PRODUCTION - PILOT PLANT (ECKENFELDER EVALUATION)

Figure 80

TABLE 32

Comparison of Predicted and Measured Values of Sludge Wasted in Full Scale Plant-Active Mass Evaluation

Period	ΔX Predicted, in lbs/day	ΔX Measured, in lbs/day
A	1514	854
B	-	-
C	-	-
D	-	-
E	-	-
F	-	-
G	-	-
H	-	-
I	3507	1067
J	3614	750
K	815	944
L	512	961
M	1182	1351
N	745	928
O	-	-

for periods 2 to 8, all values of ΔX predicted were negative, meaning no sludge wasting, when actually sludge was wasted in the routine operation.

The result of the pilot plant solids analysis is parallel to the full scale work. Namely, that more study is needed to accurately predict solids production under field conditions.

TABLE 33

Comparison of Predicted and Measured Values of Sludge Wasted in Full Scale Plant - Eckenfelder Evaluation

Period	ΔX Predicted (lb VSS/day)	ΔX Measured (lbs VSS/day)
A	606	854
B	203	690
C	661	915
D	169	2583
E	773	853
F	2480	1239
G	3498	1579
H	2970	1756
I	792	1067
J	1839	750
K	1288	944
L	1012	961
M	876	1351
N	789	928
O	394	1267

SECTION X

IMPROVED DESIGN CONCEPTS

General

The intent of some research programs is to determine with data the reliability of design concepts on full scale operations. The homogeneous mixed system has been tested extensively in the laboratory, pilot plants, and full scale industrial waste operations. The data from these different operations have indicated the potential of this activated sludge system to operate at lower detention periods and higher loading conditions (F/M and lb/BOD/1000 ft^3) than normally allowed by the state standards and still achieve the desired removal efficiencies.

An effort was made in this research program to gather data relative to the application of the homogeneous mixed system to a municipal sewage wastewater flow. The results given in other sections of this report show that the homogeneous mixed system proved to be effective in treating the Freeport wastewater flow at lower detention periods and higher loading conditions than the design criteria established by the State Agency.

The application of test results to the design of full scale installations is the desire of the design engineer. The initial design of the Freeport Plant was performed with very little previous design criteria available for a homogeneous mixed system. The completed plant design proved to be very operable but, as in any initial endeavor,some modifications in future designs are desirable. The following items describe some modifications which would result in a more optimum design.

Number of Air Diffusers Required

The original design for the air diffusion system varied for each tank due to the flexibility required for the different modes of operation. One tank had 833 diffusers, another 633 diffusers, both due to special test requirements, and the remaining two tanks each had 408 diffusers. After careful examination of the oxygen uptake data, air flow data and transfer efficiency tests given in other sections of this report, it appears that about 1,000 ft^3 of air per pound of 5-day BOD applied is needed to supply sufficient oxygen in the system. Based upon the design value of 7,900 pounds per day BOD applied to the aeration tanks, a total of 5,480 cfm is the average air demand of the system. Each fine bubble diffuser was designed to supply 6 cfm and therefore

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228 diffusers required in each tank.

The above design would satisfy the requirement for oxygen demand, but in the paper entitled "Evaluation of Mixing in Aeration Tanks" by E.D. Toerber, W.L. Paulson, S.C. Mehta, and H.S. Smith, the following statement was made: "A recommended absolute minimum density for good mixing should be 20 diffusers per 100 square feet." This criteria would result in 287 diffusers per tank.

To meet both the oxygen demand and mixing requirements of the system, each 25 foot wide by 57.5 foot long tank should be supplied with 290 diffusers resulting in a flow rate of about 5 cfm each. This would reduce the total number of diffusers in the four tanks from 2,282 to 1,160.

Another modification of the original diffuser design is the location of the diffusers in the aeration tanks. It was indicated in the dye tests (described in Section IV), that more diffusers placed on the effluent half of the aeration tank bottom resulted in more complete mixing of the tank contents. The results of this diffuser arrangement seems logical because more backmixing could improve utilization of the tank volume. Further research should be performed to confirm this point for other tank configurations.

Blower Capacity

The blower range was not adequate during some periods of plant operation. The blowers, with a range of about 3,500 to 6,000 cfm each, could not be operated at a low enough output for periods of low oxygen demand, thereby wasting some of the air produced during these periods. This problem resulted from the much lower initial flows and 5-day BOD loads that normally occur during early stages of design periods. As previously explained, only about 5,480 cfm of air is required for the aeration system. In a new design three 2,750 cfm blowers would be installed with one of the blowers as a stand-by. This arrangement would provide a range of blower output between 1,500 and 8,250 cfm, which would result in enough flexibility to cope with the different loading conditions.

Instrumentation to Regulate Blowers

One of the major deficiencies of many sewage treatment plants is lack of adequate instrumentation for optimum operation. Part of this over-all instrumentation deficiency is in the lack of automatic adjustment of blower output to meet oxygen demand. Figure 50 illustrates a typical diurnal BOD load to the Freeport sewage treatment plant.

As indicated in Figure 50, the BOD load to the treatment system will have large fluctuations during the course of the day. Therefore, in a biological treatment system the microorganisms will exert similar fluctuations in their demand for oxygen while degrading the organic material.

To properly regulate the air supply, it is customary at most plants to periodically measure the dissolved oxygen at the effluent from the aeration tanks and to adjust the blower output in an attempt to maintain the desired dissolved oxygen concentration. Recent technological advances have produced dissolved oxygen probes which can, with proper maintenance,give continuous and instantaneous readings which can be converted to control signals for automatic adjustment of blower output. This system enables the blower output to be optimized by maintaining a set dissolved oxygen level in the aeration tanks and to prevent excessive air output. Further research in this area should be pursued to provide the operators with a tool for improved plant operation.

Spacing of Inlets to Aeration Tanks

Considerable discussion has occurred concerning the proper spacing of wastewater and return sludge inlets into an aeration tank which would result in a homogeneous mixed system. Part of the present study dealt with mixing regimes existing in the aeration tanks. The mixing regimes were studied utilizing a rhodamine B dye test. The results, which are summarized in another section of this report, indicate that two inlet ports each for primary effluent and return activated sludge for each fifty feet of tank length are sufficient. The inlet ports should be located symmetrically on one side of the tank.

The inlet ports for the primary effluent exhibited a solids settling problem. Future designs should maintain enough velocity in the inlet ports to prevent excessive settling of the solids.

Optimum Detention Period in Aeration Tanks

The "Ten State Standards" state that a conventional biological treatment system should be designed for a loading of 40 pounds of 5-day BOD per 1,000 cubic feet of aeration tank volume, which would result in a detention period of 5.25 hours for the Freeport design conditions. The existing Freeport plant has exhibited good removals of soluble BOD in as little time as one hour, as shown in Figure 35.

After careful examination of the data and taking into consideration the minimum detention time required in the aeration tanks during high flow conditions, it can be stated that a two hour detention period (based upon Q) for the average design flow conditions (6.75 mgd) would be adequate for design purposes.

Economical Evaluation

General

One of the major governing factors in the design of treatment facilities is the construction cost. It is the responsibility of the engineer to provide a design which will produce the results required at the lowest possible construction cost, at the same time considering the financing capability of the client. If the engineer is able to design an optimum system without being compelled to fully comply with conservative state standards, the client would definitely profit.

The following paragraphs describe: (a) an EPA cost estimating program; (b) the estimated construction cost of the Freeport facility if designed in accordance with the criteria developed during the test program; and (c) an estimated construction cost of a treatment facility based upon the Freeport design conditions and designed in accordance with conventional design criteria.

Following the explanation of the basis for each final cost, a comparison of the three cost results is given. This comparison takes into consideration the cost of the aeration tanks, blowers, blower building, final sedimentation tanks, return sludge pumps, associated piping, plumbing, heating, ventilation, air conditioning, equipment, excavation, electrical work, roadways and landscaping. The remainder of the treatment facility units (primary treatment, sludge disposal, disinfection) were not included because the units would be similar for all designs.

EPA Cost Estimating Program

The U.S. Environmental Protection Agency has devised a computer program (Wastewater Treatment Plant Cost Estimating Program, by R. Eilers and R. Smith) which estimates the construction costs of treatment plants. The basic data utilized by the EPA in creating the computer program was based upon past construction cost records kept by a major consulting engineering firm. These costs were broken down into the cost for each major process of the over-all treatment facility, namely, aeration tanks, final sedimentation tanks, etc. The construction cost for each process is related to flow, surface overflow rate, liquid volume, BOD_5 concentration, etc., depending upon the major process under consideration. For example, the cost of the final sedimentation tanks is related to the surface overflow rate.

The final tank loading deserves special explanation. The surface overflow rate chosen for purposes of comparison was 800 gpd/ft^2. This was based on an engineering judgement by Consoer, Townsend and Associates who did the design analysis. This differs from the loadings recommended in Section VII (Surface Overflow Rate = 700 gpd/ft^2, Solids Loading = 25 lb/ft^2·day). The comparison does not take into account the relatively low solids loading recommended. The type of final settling tank will also affect the choice of allowable loadings.

Based upon the above criteria, the following parameters were inserted into the program:

BOD_5 Removal in Aeration Tanks	125 mg/l
MLVSS	2,000 mg/l
Influent Dissolved Nitrogen Concentration	15 mg/l
RSVSS/MLVSS Ratio	4.2
Aeration Detention Time	2.0 hours
Diffuser Efficiency	8%
Final Sedimentation Tank Overflow Rate	800 gpd/ft^2
Average Daily Flow	6.75 mgd
Peak Flow	16.6 mgd

The computer program listed the following costs for each unit:

Aeration Basin Structure	$ 99,126
Aeration Diffused Air	332,208
Final Sedimentation (Multiple Basin)	207,759
Recirculation Pumping	94,339
Yardwork	102,683
Total Estimated Construction Cost...$	836,115

The total construction cost is based upon an ENR Index of 1308.61.

Estimated Construction Cost, Homogeneous Plant

The actual construction costs of the following listed existing units at the Freeport Plant were determined when the Engineering News-Record Construction Cost Index was about 1140.0:

Tunnel	$ 51,000
Blower Building	508,000
Aeration Tanks	328,000
Final Tanks	334,000
Outside and Miscellaneous Work	100,000
Total Construction Cost	$1,321,000

Taking into consideration the design modifications previously described in this section of the report, namely, size of the blowers, number of inlets, number of diffusers and the size of the final tanks, the cost of the following listed units would be modified as shown:

Blower Building	$ 340,000
Aeration Tanks	258,000
Final Tanks	300,000

To the above values add the previous tunnel and miscellaneous work cost, and the resulting adjusted construction cost is $1,049,000 at an ENR Index of 1140.0.

To have the same basis for comparison with the other construction costs, the January, 1970 ENR Index of 1308.61 was chosen and the resulting construction cost is $1,200,000.

Estimated Construction Cost, Conventional Design Plant

The "Ten State Standards" have established criteria for design of sewage treatment facilities. Utilizing part of the above criteria in designing a conventional biological system, a construction cost estimate of the final design was made. The criteria which affect the biological system are as follows:

(a) 40 pounds of 5-day BOD per 1,000 cubic feet of aeration capacity

(b) 1,500 cubic feet of air per pound of 5-day BOD to the aeration tank

(c) 75% return rate capacity of the average design flow

(d) 800 gallons per day per square foot of final tank surface area

The average design flow is 6.75 mgd and the BOD load to the aeration tanks is 7,900 pounds per day, as previously stated in this section.

3 OF 3

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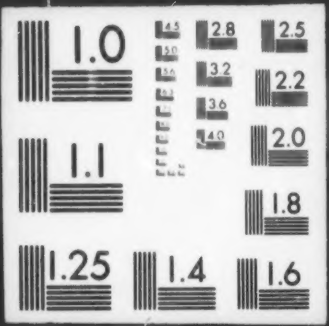
1.0
4.5
5.0
5.6
6.3
2.8
3.2
3.6
4.0
2.5
2.2
2.0
1.1
1.8
1.25
1.4
1.6

Based upon the above criteria and design conditions, the following capacities would be required:

(a)	Aeration Tanks	197,000 cu ft
(b)	Blower Capacity	8,200 cfm
	Use 3 - 4, 120 cfm blowers with one as stand-by	
(c)	Diffusers, @ 6cfm/diffuser	1,370
(d)	Return Activated Sludge Pump Capacity	5.1 mgd
	Use 4 - 1.3 mgd variable speed pumps	
(e)	Final Tanks	8,430 sq ft

The comparative cost associated with the equipment and tankage for the above items based upon an ENR Index of 1308.61 is as follows:

Tunnel	$ 60,000
Blower Building	520,000
Aeration Tanks	950,000
Final Tanks	450,000
Outside and Miscellaneous Work	300,000
Total Construction Cost	$2,280,000

Comparison of Different Construction Costs

The following table lists the construction costs associated with the three different economical evaluations of the treatment facilities based upon the previously described design criteria. All the costs are based upon an ENR Index of 1308.61.

Economical Evaluation Method

EPA Cost Estimating Program	$ 836,115
Estimated Construction Cost, Homogeneous Plant	1,200,000
Estimated Construction Cost, Conventional Design Plant	2,280,000

The standard to be used in the comparison is the estimated construction cost of the facilities if designed in accordance with the criteria developed during the test program. The EPA cost estimating program was considerably lower than the estimated construction cost.

It should be noted that the EPA program estimate can vary by as much as ± 40% of the actual construction cost of a particular plant, due to the many special considerations that may arise in building a new plant. Cost variations are due to differences of design, local construction site conditions, current and projected labor costs, climate and seasonal factors, the economic condition at the time construction bids are taken, and many other factors which are not readily subject to evaluation on an average basis. The EPA program is only intended to provide a preliminary construction cost estimate for an average plant (strictly speaking, no average plant really exists). Therefore, the comparison between the two costs is within the intended range of the computer program estimate.

This research program has proved the Freeport homogeneous activated sludge plant to be an efficient wastewater treatment facility which has cost the Freeport residents only about one-half as much as a conventional plant as indicated in the comparison previously given. The Freeport demonstration also strongly indicates that more latitude should be allowed the designer in the application of design standards established by state agencies or that the standards for activated sludge treatment should be modified.

It is the intent of the engineer to design a facility that satisfies the conditions at the lowest cost. This approach will challenge the engineer's innovative ability because many aspects must be considered in arriving at the final design. If the engineer is hampered by too conservative criteria, part of the incentive and challenge is removed. Not only is the engineer affected adversely, but the overall process efficiency suffers and the taxpayer bears the added cost. Treatment efficiencies have suffered in some cases because over-designed facilities do not promote maximum performance. Also the taxpayer is required to pay for extra facilities which are not needed to obtain the desired results.

It is hoped that the results of this research program will lead to more efficiently designed biological wastewater treatment plants.

EPA Process Performance Simulations

The EPA has developed another computer program (A generalized Computer Model for Steady-State Performance of the Activated Sludge Process, by R. Smith and R. Eilers). This program is a steady-state model of the conventional activated sludge process which can be used to simulate all performance aspects of the process. This model has been shown to fit measurement data from a wide range of process models, including complete-mix, plug-flow, step aeration, contact stabilization, extended aeration, etc.

This program was used to simulate some of the measurement

COMPARISON OF ACTIVATED SLUDGE MEASUREMENTS AND MODEL COMPUTATIONS

TABLE 34

	Complete-Mix Mode 4-24/5-9, 1971		Complete-Mix Mode 1-1/1-31, 1971		Parallel Operation 7-1/7-31, 1971			
					Complete-Mix Mode		Plug-Flow Mode	
	M	C	M	C	M	C	M	C
MLVSS, mg/1	3301	3309	2501	2490	1166	1192	1143	1177
TBOD, mg/1	33	30	19	18	31	16	28	13
DBOB, mg/1	14	20	14	12	16	12	15	6
NH3, mg/1	16	16	20	20	14	14	14	14
Q15, mgd	.02	.03	.01	.03	.03	.06	.02	.05
Q17, mgd	.90	.96	.70	.60	.62	.52	.61	.37
AIR, scfm	-	-	-	-	1083	830	1539	885

	Parallel Operation 8-1/8-31, 1971				Parallel Operation 9-1/9-30, 1971			
	Complete-Mix Mode		Plug-Flow Mode		Complete-Mix Mode		Plug Flow Mode	
	M	C	M	C	M	C	M	C
MLVSS, mg/1	1333	1321	1230	1244	2111	2125	1795	1799
TBOD, mg/1	17	12	16	9	19	13	17	8
DBOD, mg/1	8	7	8	4	9	8	11	3
NH3, mg/1	14	14	14	14	14	14	14	14
Q15, mgd	.03	.05	.02	.05	.02	.03	.01	.03
Q17, mgd	.65	.65	.64	.57	.74	.57	.66	.62
AIR, scfm	996	625	1551	657	1343	772	1477	792

Note: Refer to page 186 for Key to all abbreviations

data which was gathered on this project. A summary of this effort appears in Table 34. Based on raw data averages over a selected simulation period, several measured system parameters are chosen for comparison to predicted values from the computer program. Note that several simulations of the parallel operation data are given in Table 34. Based on these results, it can be concluded that the model can produce a fairly good simulation of the complete-mix mode of operation, but the program consistently predicts that better performance should have been obtained from the plug-flow mode; which may indicate a deficiency in the activated sludge computer model.

Below is a tabulation of the abbreviations used in Table 34:

MLVSS = mixed liquor volatile suspended solids in the aerator, mg/l
TBOD = final effluent unfiltered 5-day BOD, mg/l
DBOD = final effluent filtered 5-day BOD, mg/l
NH3 = final effluent ammonia nitrogen, mg/l
Q15 = waste stream flow from the activated sludge process, mgd
Q17 = return stream flow to the activated sludge process, mgd
AIR = total minimum air requirement for the activated sludge process, scfm
M = average measured value from the raw data
C = computed value from the program

SECTION XI

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Thanks must first go to Mr. Lawrence M. Madden who served as project manager. His enthusiastic support of the project engineer should be recognized. Along with Mr. Madden, the Freeport Water and Sewer Commission are to be commended for their farsightedness in authorizing construction of a plant with enough flexibility to be adapted to the research study.

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The research consultants, Dr. H.S. Smith and Dr. Wayne L. Paulson, provided the majority of the ideas for the plan of study conducted, reviewed the final report and made numerous revisions to improve this report.

The operating personnel at the Freeport Wastewater Treatment Plant should be acknowledged for the analytical and operational work that they performed. The following were the key persons:

Mr. Joel McCulloch, chemist
Mr. Glenn Kelly, chief operator
Mr. Michael Ross, lab technician
Mr. Ralph Sager, technician
Mr. Cy Crothers, maintenance operator
Mr. Donald Maher, maintenance operator

Mr. Suman Mehta was in charge of the mixing study phase of the project.

Mr. Bharat Mathur was in charge of the pilot plant study and wrote this section of the final report.

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SECTION XII

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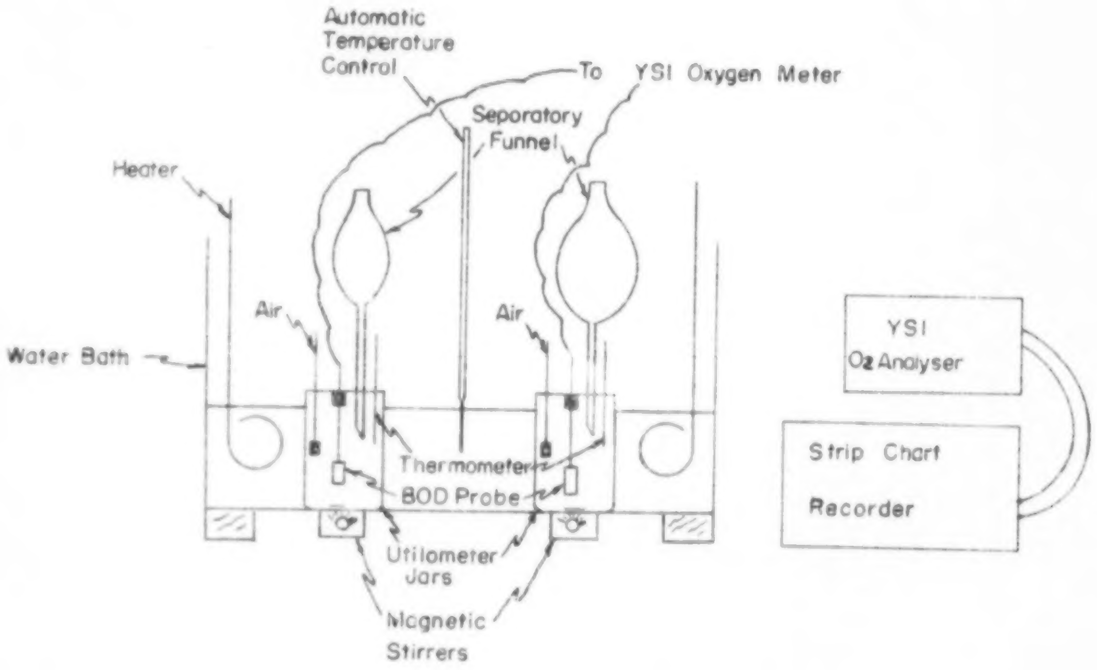
SECTION XIII

APPENDIX

Oxygen Uptake Tests

A schematic diagram of the oxygen uptake test apparatus shown in Figure 81 (This figure is appropriately labelled and indicates the apparatus involved). Initially, the YSI oxygen probe is calibrated. In this investigation it was calibrated against final effluent. The following steps are carried out:

1. Pour 1100 cc tap water in a utilometer and aerate.
2. Attach separatory funnel to utilometer.
3. When sufficiently high DO is read on the recorder, start stop watch and collect sample (mixed liquor in this case). This is stop watch zero (t_1).
4. The time at which the sample is withdrawn from the aeration tank is noted (t_2). This is zero for the sample (mixed liquor in this case).
5. Pour 1500 ml sample into the utilometer. Save a portion for solids determination.
6. Seal the utilometer and note the temperature.
7. Allow oxygen consumption until DO drops below 1 mg/1.
8. Reaerate to a sufficiently high DO and reseal.
9. Repeat 7 and 8 and run experiment for 24 hours.


Automatic Temperature Control
To YSI Oxygen Meter
Seporatory Funnel
Heater
Air
Air
Water Bath
Thermometer
BOD Probe
Utilometer Jars
Magnetic Stirrers
YSI O2 Analyser
Strip Chart Recorder

OXYGEN UPTAKE TEST APPARATUS
Figure 81

Calculation of Uptake

From the strip chart the following parameters are recorded:

1. Slope of line between time interval t' - t"
2. Time interval t' - t"
3. Time $t = t' + \frac{t' - t''}{2}$

Temperature and calibration corrections are applied to give the uptake in mg/l/hr at 20° C.

A curve is plotted between oxygen uptake do/dt in mg/l/hr and t. Refer to Figure 82 for a typical curve. The straight line portion of the curve is extended to meet t = 0, in this case at do/dt = 13.5 mg/l/hr. The slope of the straight line is determined as follows:

$t = 0$, for $do/dt = 13.5$

$t = 10$ hours, for $do/dt = 9.9$

then $K_2 = \frac{\log 9.9 - \log 13.5}{10} = 0.029 \text{hrs}^{-1}$

then M_a in oxygen units $= \frac{13.5}{.0296} = 450$ mg/l

or M_a in mass units $= \frac{450}{1.42} = 317$ mg/l

Since VSS was determined to be 1190 mg/l

$$\frac{M_a}{VSS} = 0.26$$

The difference between the curved portion of the plot and the straight line is then plotted for a convenient time interval. The slope of this line is determined as before to be = 2.67 hrs^{-1}, which is K_9. Then $\frac{7.7}{K_9}$ = F in mg/l. **F is ultimate BOD.**

Oxygen Transfer Efficiency Test

This test was conducted by using the pilot plant aeration tank. The following steps detail the experimental procedure:

1. Fill tank with sample in which transfer efficiency is desired and choose air flow rate.
2. Calibrate DO probe and immerse in sample to get continuous DO monitoring.
3. Record the DO concentration at saturation, C_s and also the liquid temperature.
4. Introduce a pre-determined amount of sodium sulphite mixed with cobalt chloride catalyst (to give a C_oCl_2 concentration in the tank of 1.5 mg/l). Start the stop watch. This is time zero. If appropriate

quantities of sulphite have been added the DO should go to zero.

5. When the DO begins to increase, record the time in minutes and C_t, the DO corresponding to that time, until DO reaches 80% of saturation value.

The following steps are involved in the calculation of the transfer efficiency:

1. Plot $C_s/(C_s - C_t)$ versus time on a semi-log paper.
2. Find slope of curve = K at temperature of liquid, say K_T.
3. Correct K_T for 20° C, where $K_{20} = K_T(1.024)^{20-T}$
   Then rate of oxygen consumption $\frac{dc}{dt} = C_s \times K_{20} \times 60$ mg/l/hr.
4. Convert air supply rate to mg/l/hr. Then transfer efficiency $= \frac{\text{rate of oxygen consumption}}{\text{rate of air supply}}$

   α, the oxygen transfer coefficient $= \frac{K(\text{waste water})}{K(\text{tap water})}$

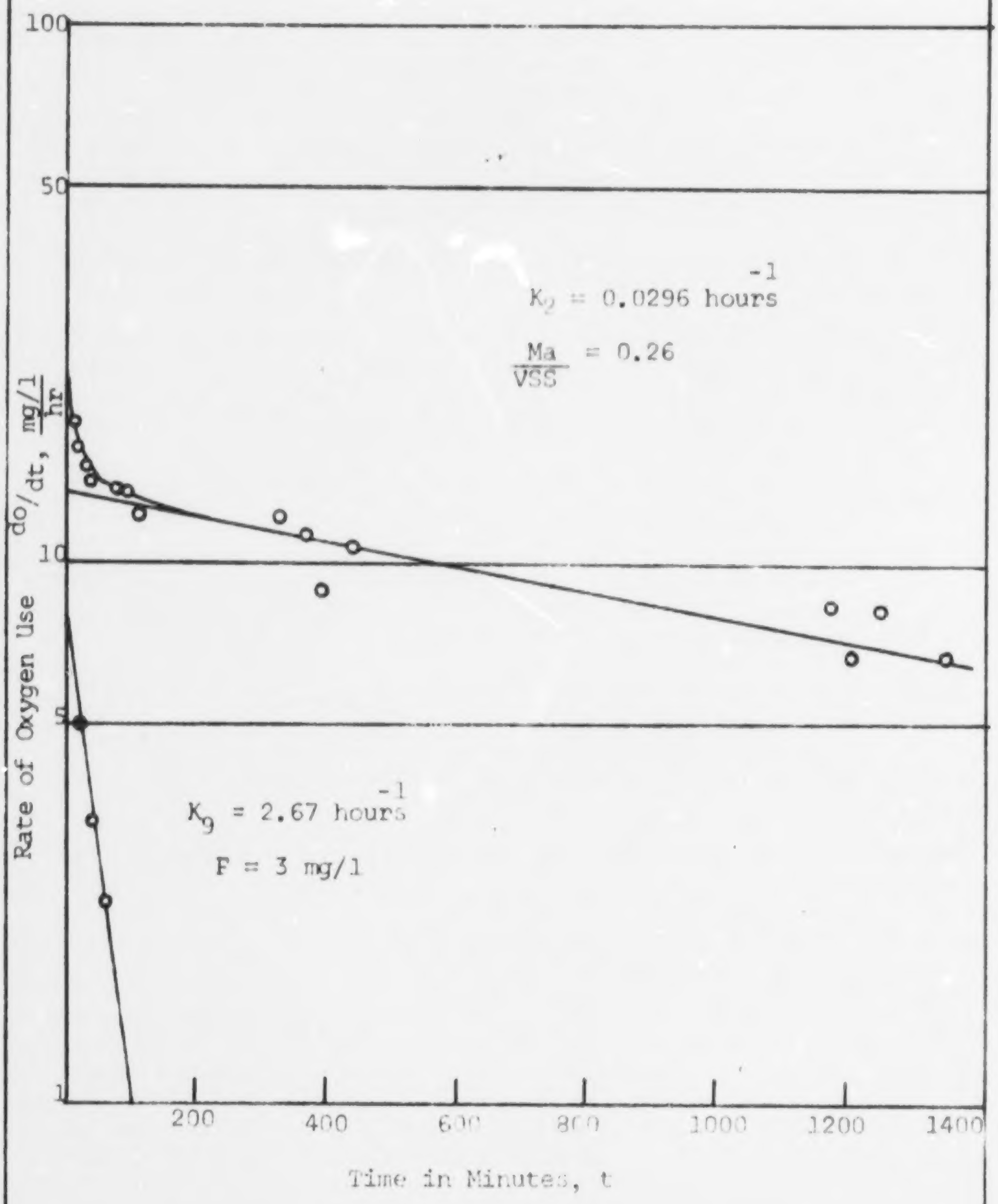


Figure **82** Oxygen uptake, Run No. 7/1/PP

Biochemical Oxygen Demand Test

1. Apparatus

1.1 Incubation bottles, 300 ml capacity with ground glass stoppers. Bottles should be cleaned with a good detergent and thoroughly rinsed and drained before use. The BOD bottles should have flared mouths so that an air tight water seal can be provided during incubation.

1.2 Air incubator thermostatically controlled at 20° C ± 1° C. All light should be excluded to prevent formation of DO by algae in the sample.

1.3 Yellow Springs Instrument Company, Model 54 oxygen meter.

1.4 Y.S.I. Model S420A self stirring BOD bottle oxygen probe. The probe is especially designed for use with standard BOD bottles. The probe is provided with an agitator for stirring of the sample solution at a constant rate.

2. Reagents

2.1 Distilled water for preparation of dilution water. This water must contain less than 0.01 mg/l copper and be free of chlorine, choloramines, caustic alkalinity, organic materials or acids.

2.2 Phosphate buffer solution: Standard Methods - 12th Edition

2.3 Magnesium sulfate solution: Standard Methods - 12th Edition.

2.4 Calcium chloride solution: Standard Methods - 12th Edition.

2.5 Ferric chloride solution: Standard Methods - 12th Edition.

2.6 Seeding material: 2 liters of mixed liquor from the aeration tank should be placed into a glass container and continually aerated. Daily 500 ml of mixed liquor should be drawn from the aeration container. Place mixed liquor into 1 liter graduate cylinder and allow solids to settle. 10 ml of clear liquor should then be drawn off this solution and diluted with distilled water to 400 ml. The 500 ml of mixed liquor should be replaced by adding 500 ml of primary effluent to the aeration container. A solid level of no less than 100 ml/l should be maintained in the aeration container. This level may be maintained by adding return sludge or mixed liquor to replace the 500 ml portion that was drawn from the aeration container.

3. Procedure

3.1 Preparation of dilution water: The distilled water that is used is aerated with clean compressed air for a sufficient length of time to become saturated with DO. The distilled water should be as near 20° C as possible. The distilled water is placed in a 2.5 gallon bottle and 10 ml each of phosphate buffer, magnesium sulfate, calcium chloride and ferric chloride solutions are added .

3.2 Seeding: 2 ml of seed as prepared in paragraph 2.6 should be added to each BOD bottle.

3.3 Pre-treatment of sample:

a) Pasteurization: samples should be heated to 55° C for 15 minutes and allowed to cool to room temperature before use (only final effluent samples are pasteurized).
b) Filtering: filtered samples should be filtered through a glass fiber filter (Gelamn Type A, part 61693). For comparison, both filtered and unfiltered samples are run simultaneously.

3.4 Dilution Technique: raw sewage is diluted by placing 10 ml of sample by pipette into a BOD bottle half filled with dilution water. The same dilution procedure is used for primary effluent (filtered and unfiltered). Final effluent unfiltered is diluted by placing 25 ml of sample by pipette into a BOD bottle half filled with dilution water and then filling the BOD bottle to the top. 50 ml of sample is used for final effluent filtered.

4. Calibration of Oxygen Meter and Probe

4.1 Two BOD bottles should be filled with dilution water. The water should be added by a siphon with it's outlet held below water level to prevent further aeration.

4.2 The DO of the first bottle should then be determined by use of the Winkler test for dissolved oxygen.

4.3 Place the YSI probe into the second BOD bottle and calibrate meter to the DO reading found by the results of the Winkler test of the first BOD bottle.

4.4 A Winkler test should then be run on the second BOD bottle. If and only if the results of the two Winkler tests are the same, are the YSI probe and meter accurately calibrated.

5. Use of Oxygen Meter and Probe

5.1 Initial DO: BOD bottle should be prepared with seed and sample as described in paragraphs 3.2,3.3, and 3.4. The probe should be inserted into bottle and a DO reading obtained. The sample is then incubated in the BOD bottle for five days at 20°C.

5.2 Final DO: After five days of incubation the BOD bottle should be removed from the incubator, and the DO determined by the YSI meter and probe.

6. Calculation of BOD

$$BOD = (I-F) \times \frac{300}{P}$$

I = Initial DO
F = Final DO
P = Sample size in ml

A blank is run with each bottle of dilution water to determine any depletion in DO of the dilution water.

Chemical Oxygen Demand Test

1. Apparatus

1.1 Refluxing apparatus, 250 ml erlenmeyer flasks with ground glass 24/40 neck and 300 mm jacket, liebeg condensers with 24/40 ground glass joint, electric hot plate extraction heater (6 burner) producing 10.3 watts/sq in of heating surface to provide adequate boiling of contents of the refluxing flask.

2. Reagents

2.1 Stock potassium dichromate 0.250 N with 0.129 of sulfamic acid added to prevent nitrite interference.

2.2 Standard potassium dichromate solution 0.10N made by diluting 400 ml of stock to 1000 ml.

2.3 Sulfuric acid reagent, conc. H_2SO_4 containing 22 g silver sulfate $Ag_2 SO_4$, per 9-lb bottle.

2.4 Stock ferrous ammonium sulfate solution 0.10N. Dissolve 39g $Fe(NH_4)_2 (SO_4)_2 \cdot 6H_2O$ in distilled water. Add 20 ml conc. H_2SO_4, cool, and dilute to 1000 ml.

2.5 Standard ferrous ammonium sulfate titrant 0.04N. Made by diluting 400 ml of stock solution to 1000 ml. This solution must be standarized daily against the standard potassium dichromatic solution.
Standardization: Dilute 10.0 ml standard potassium dichromate solution to about 100 ml. Add 30 ml conc. H_2SO_4 and allow to cool. Titrate with the ferrous ammonium sulfate titrant, using 3 drops of ferroin indicator.

$$N = \frac{\text{ml } K_2 Cr_2 O_7 \times 0.25}{\text{ml } Fe(NH_4)_2 (SO_4)_2}$$

2.6 Ferroin indicator solution: Fisher Scientific Co. # P-69, 1, 10 Phenanthroline ferrous sulfate complex.

2.7 Silver sulfate reagent powder.

2.8 Mercuric sulfate, analytical grade crystals.

2.9 Sulfamic acid, analytical grade.

3. Procedure

3.1 To each glass refluxing flask (blank and sample flasks) add 0.4g mercuric sulfate. This is measured with a Hach Company #638 measuring spoon.

3.2 Add 20.0 ml boiled distilled water to the blank and 20.0 ml of sample to the sample flasks. Filtered and unfiltered sample are used. Filtered samples are filtered through a Gelman Type A glass fiber filter, Part #61693.

3.3 Add exactly 10.0 ml standard potassium dichromate 0.10 N solution to each flask.

3.4 Carefully add to each flask 30 ml of the concentrated sulfuric acid containing the dissolved silver sulfate and thoroughly mix by swirling.

3.5 Add 10 to 15 3mm glass boiling beads.

4. Attach condenser and reflux both blank and sample for 2 hours.

5. Cool and wash down the condenser with 80 ml of distilled water. This should give a total volume of about 140 ml. This volume is necessary to allow the indicator to function properly.

6. After sample is cooled to room temperature and diluted, add 2 or 3 drops of ferrion indicator to each flask. Use the same number of drops for each flask.

7. Titrate each flask with the standarized ferrous ammonium sulfate. The color change is sharp, going from blue-green to reddish-brown to indicate the end point.

8. Calculation:

$$\text{COD, mg/l} = \frac{(a-b)c \times 8{,}000}{\text{ml of sample}}$$

a = ml ferrous ammonium sulfate solution used for blank
b = ml ferrous ammonium sulfate solution used for blank
c = Normality of ferrous ammonium sulfate solution

Mixed Liquor Suspended Solids & Volatile Suspended Solids Test

1. Apparatus

   1.1 Gooch crucible

   1.2 Gelman #61693 glass fiber filter, type A, 1 inch

   1.3 Vacuum filter flask

2. Procedure

   2.1 A sample of 1 liter volume should be taken

   2.2 The crucible and glass fiber filter should be oven dried at 103° C to 105°C for one hour. Allow to cool in desicator and weigh.

   2.3 Sample should be thoroughly mixed and 10 ml of sample placed in tared crucible under vacuum. Leaving vacuum applied, rinse the measuring pipette with 10 ml distilled water and filter through glass fiber filter.

   2.4 Dry crucible and solids for two hours at 105°C. Allow to cool in desiccator and weigh.

   2.5 For volatile solids, ignite the crucible for 15 minutes at 600°C in a muffle furnace. Cool to room temperature, partially in air and finally in a desiccator, and weigh.

3. Calculation:

   mg/liter total suspended solids =

   $$\frac{\text{mg weight of dried solids} \times 1000}{\text{ml of sample}}$$

   mg/liter volatile suspended solids =

   $$\frac{\text{mg weight of volatile solids} \times 1000}{\text{ml of sample}}$$

   The volatile solids are equal to the dried total solids minus the residue upon ignition.

4. Reporting

   Report as mg/l suspended solids and mg/l volatile suspended solids.

Ammonia Nitrogen Test

1. Apparatus

   1.1 1000 ml Florence flask, used for distillation

   1.2 Vertical condenser (glass)

   1.3 Glass tubing, stoppers

   1.4 Electric heater

   1.5 500 ml Erlenmeyer flask, used for sample collection

   1.6 Spectrophotometer for use at 425 nm and providing a lightpatch of 1 cm or more

   1.7 Glass boiling beads

   1.8 Nessler tubes

   1.9 5 ml and 20 ml volumetric pipettes

2. Reagents

   2.1 Ammonia free distilled water

   2.2 Rochelle salt solution

   2.3 Stock ammonium solution, 1.00 ml = 1.00 mgN

   2.4 Standard ammonium solution, 1.00 ml = 0.01 mgN

   2.5 Nessler reagent (Fisher SO-N-16)

   2.6 Phosphate buffer solution

3. Procedure

   3.1 400 ml of sample should be placed in Florence flask (1000 ml) and 25 ml of phosphate buffer solution added to the sample.

   3.2 Distillation: distill 200 ml at the rate of 3-5 ml/min into 500 ml Erlenmeyer flask.

   3.3 Pipette 5 ml of distillate into 50 ml Nessler tube. Dilute sample to 50 ml with ammonia free distilled water, add two drops of Rochelle salt and 1 ml of prepared Nessler reagent.

3.4 Color Developement: Allow 10 minutes for color developement. Read the percent transmittance at 425 nm against the blank. From values obtained, find concentration (mg/l) on standard curve, transmittance vs. concentration.

3.4-1 Standard curve preparation. Prepare a series of standards in the following concentrations: 0.0, 0.1, 0.3, 0.5, 0.8, 1.0 mg/l. Nesslerize standards and allow 10 minutes for color development. Read transmittance at 425 nm. From values obtained, plot percent transmittance against concentration for the standard curve.

4. Calculation

$$\text{Ammonia N, mg/l} = \frac{(50)\ A}{\text{ml sample}} \times \frac{B}{C}$$

A = mg/l N found colorimetrically

B = ml total distillate collected

C = ml distillate taken for nesslerization

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